PB80-199813

Proceedings: The Fifth Industry Briefing on IERL-RTP Lime/Limestone Wet Scrubbing Test Programs (December 1979)

(U.S.) Industrial Environmental Research Lab.
Research Triangle Park, NC

Jun 80

U.S. Department of Commerce
National Technical Information Service
NTIS

600/9-80-032
June 1980

PB80-199813

EPA

PROCEEDINGS: FIFTH INDUSTRY BRIEFING ON IERL-RTP LIME/LIMESTONE WET SCRUBBING TEST PROGRAMS
(DECEMBER 1979)

John E. Williams, Conference Chairman

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, North Carolina 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, D.C. 20460

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

NOTICE

THIS DOCUMENT HAS BEEN REPRODUCED FROM THE BEST COPY FURNISHED US BY THE SPONSORING AGENCY. ALTHOUGH IT IS RECOGNIZED THAT CERTAIN PORTIONS ARE ILLEGIBLE, IT IS BEING RELEASED IN THE INTEREST OF MAKING AVAILABLE AS MUCH INFORMATION AS POSSIBLE.

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO.
EPA-600/8-80-032

2.
IERL-RTP-1059

3. RECIPIENT'S ACCESSION NO.
PB 80 199813

4. TITLE AND SUBTITLE
Proceedings: Fifth Industry Briefing on EPA Lime/ Limestone Wet Scrubbing Test Programs (December 1979)

5. REPORT DATE
JUNE 1980 ISSUING DATE

IZATION CODE

7. AUTHOR(S)
John E. Williams, Conference Chairman

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS
See Block 12

10. PROGRAM ELEMENT NO.
INE828

11. CONTRACT/GRANT NO.
N.A. (Inhouse)

12. SPONSORING AGENCY NAME AND ADDRESS
EPA, Office of Research and Development
Industrial Environmental Research Laboratory
Research Triangle Park, NC 27711

13. TYPE OF REPORT AND PERIOD COVERED
Proceedings; 12/5/79

14. SPONSORING AGENCY CODE
EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is John E. Williams, MD-61, 919/541-2483. EPA-600/7-79-092 is the proceedings of the August 1978 conference.

16. ABSTRACT The proceedings document presentations made during the December 5, 1979, industry briefing conference which dealt with the status of EPA/IERL-RTP's flue gas desulfurization (FGD) research, development, and application programs. Subjects considered included: lime/limestone scrubbing test results involving the organic buffer, adipic acid; limestone type and grind tests at IERL-RTP's pilot plant; IERL-RTP's program on waste solids disposal; the TVA/Shawnee lime/ limestone design economic study computer program; alkali feedrate automatic control for limestone scrubbers; tests on the cocurrent scrubber at TVA/Shawnee; and forced-oxidation tests at the Widow's Creek Unit 8 full-scale limestone scrubber. The conference provided developers, vendors, users, and those concerned with regulatory guidelines with a current review of progress made in IERL-RTP's FGD technology development program.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Oxidation	Pollution Control	13B	07C
Flue Gases	Waste Disposal	Stationary Sources	21B	
Sulfur Oxides	Power	Forced Oxidation	07B	14G
Desulfurization	Operating Costs	Energy Requirements	07A,07D	14A,05A
Scrubbers	Adipic Acid		13I	
Calcium Oxides				
Calcium Carbonates				

18. DISTRIBUTION STATEMENT
Release to Public

19. SECURITY CLASS *(This Report)*
Unclassified

20. SECURITY CLASS *(This page)*
Unclassified

21. NO. OF PAGES

22. PRICE

EPA Form 2220-1 (9-73)

PREFACE

More than half of all "man-made" sulfur dioxide (SO_2) is emitted by electric power plants, and the use of sulfur-containing fossil fuels, especially coal, to generate electricity is expected to increase dramatically in the next 10 years. Therefore, the development and commercial application of SO_2 control technologies is one of the most important concerns of the U.S. Environmental Protection Agency (EPA). Flue gas desulfurization (FGD) is the most promising technique for control of SO_2 that will be available for widespread application to fossil fuel-fired electric power plants for at least the next 8 to 10 years.

The Industrial Environmental Research Laboratory - Research Triangle Park (IERL-RTP) of EPA's Office of Research and Development periodically sponsors symposia and Industry Briefing Conferences for the transfer of information regarding FGD research, development and application activities with the objective of further accelerating the development and commercialization of this technology. One of the major IERL-RTP FGD efforts for the past several years has been advancement of the technology for lime/limestone wet scrubbing. The focal point of this program has been the prototype testing at EPA's Alkali Wet Scrubbing Test Facility, located at TVA's Shawnee Steam Plant, near Paducah, Kentucky. Current emphasis of the test program at Shawnee is to optimize lime and limestone systems in the areas of improved sludge disposal, performance, reliability, and process economics.

The December 1979 Fifth Industry Briefing Conference focused primarily on recent test results at Shawnee in which small quantities of an organic acid buffer, adipic acid, is added to limestone/lime scrubbers to enhance SO_2 removal efficiency, improve limestone utilization, and improve control and operability of these systems. Tests were made both with and without forcing oxidation of the predominantly calcium sulfite reaction products to calcium sulfate (gypsum). Potential advantages of using adipic acid compared to other additives were also discussed. Other material presented during the conference included: results of limestone type and grind tests at the IERL-RTP pilot plant; a review of the IERL-RTP program in the waste solids disposal area; a review of the current status of the Shawnee lime/limestone design economic study computer program; a report on the investigation of alkali feedrate automatic control for limestone scrubbers; and a summary of results of TVA testing on the cocurrent scrubber at Shawnee and of the forced oxidation tests at the Widow's Creek Unit 8 full-scale limestone FGD system.

More than 100 people representing electric utilities, process suppliers, and State and Federal regulatory agencies, attended the Conference. The Chairman of the December 1979 Industry Briefing Conference was John E. Williams, a Chemical Engineer in the Emissions/Effluent Technology Branch, IERL-RTP.

These Proceedings are comprised of copies of the participating authors' papers as received.

CONTENTS

Page

LIMESTONE TYPE-AND-GRIND TESTS AT EPA/IERL-RTP 1
R. H. Borgwardt
T. J. Campbell

TEST RESULTS ON ADIPIC ACID-ENHANCED LIME/LIMESTONE SCRUBBING AT THE EPA SHAWNEE TEST FACILITY - SECOND REPORT 27
D. A. Burbank
S. C. Wang

TVA/EPRI SHAWNEE COCURRENT SCRUBBER TEST RESULTS 115
L. J. Henson

THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM 167
C. D. Stephenson
R. L. Torstrick

INVESTIGATION OF ALKALI FEEDRATE AUTOMATIC CONTROL FOR THE SLURRY PROCESS OF LIMESTONE SCRUBBERS 223
Patrick Garrett

FLUE GAS DESULFURIZATION SLUDGE DISPOSAL - THE SHAWNEE FIELD EVALUATION PROJECT 243
P. R. Hurt
P. P. Leo
Jerome Rossoff
J. R. Witz

DEMONSTRATION OF FORCED OXIDATION AT TVA'S WIDOWS CREEK UNIT 8 LIMESTONE FGD SYSTEM 275
R. A. Runyan

LIMESTONE TYPE-AND-GRIND TESTS AT EPA/IERL-RTP

by

R. H. Borgwardt
T. J. Campbell

Emissions/Effluent Technology Branch
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

and

J. H. Dempsey
J. H. Laughlin

Acurex Corporation
Mountain View, California

Abstract

Three types of high-calcium limestones were tested in the pilot TCA scrubber to compare SO_2 removal as a function of grind and stoichiometric ratio. The results confirm that both type and grind can have significant effects on TCA performance. At a stoichiometric ratio of 1.5 fine Stone Man limestone removed 74 percent of the SO_2, which was marginally better than Georgia marble (70 percent), but significantly less than that obtained with fine Fredonia (88 percent) at the same stoichiometric ratio. To maintain a given SO_2 removal, the coarse grinds (70 percent - 200 mesh) required about 50 percent greater limestone feed rates than were needed with fine (84 percent - 325 mesh) grinds. The settling rate and filterability of the sludge were also affected by the type of limestone used.

For Presentation at:

5th Shawnee Industry Briefing Conference
Raleigh, NC
December 5, 1979

INTRODUCTION

Pilot plant studies conducted by Shah (1) in 1971 showed significant differences in SO_2 removal in a two-stage venturi scrubber when testing 14 different types of limestones. Particle size was also very important, the various stones averaging about 20 percent higher SO_2 removal with a 96 percent - 325 mesh grind than was obtained with a grind of 98 percent - 200 mesh. Laboratory studies by Drehmel (2) reported marked differences in the dissolution rates of various types of stone, with Georgia marble consistently showing the lowest reactivity of seven non-dolomitic types compared by three different methods of measurement. Some preliminary type-and-grind tests at Shawnee have indicated a slightly higher SO_2 removal in the TCA (3) with Longview (Alabama) limestone, and lower removal with Stone Man (Tennessee) limestone in the venturi/spray tower (4) when compared with the Fredonia White limestone normally used in those scrubbers. Fineness of grind appeared to have the greatest effect on SO_2 removal efficiency when compared to the effects of other variables; e.g., hold tank design.

The importance of limestone grind is emphasized by a recent TVA economic evaluation (5) that considered the effect of the higher stoichiometric ratios required for a given SO_2 removal when using coarsely ground stones. Assuming stoichiometric ratios of 1.15, 1.35, and 1.55 to represent the limestone feed requirements when using grinds of 95 percent - 325 mesh, 90 percent - 200 mesh, and 70 percent - 200 mesh, respectively, the TVA study concluded that the economics favor fine grinding. Thus, the higher capital investment and power needed for the grinding equipment used to produce smaller limestone particles is more than offset by the lower costs for waste disposal and raw limestone makeup.

The Stone Man limestone is of particular interest because it is being used in TVA's full-scale flue gas desulfurization (FGD) system at Widows Creek. The Shawnee tests, which simulated the Widows Creek system, indicated that a stoichiometric ratio of 1.3 was required for that stone to equal the SO_2 removal obtained with Fredonia limestone at a stoichiometric ratio of 1.2 (6). That comparison was somewhat obscured by the use of different coals, different scrubbing configuration, and somewhat different grind than is normally used in Shawnee testing. Stone Man was therefore selected as one of the three limestone types examined at RTP; Fredonia and Georgia marble were the other two.

Prior to this time, the only limestone type (other than Fredonia) tested at RTP was Bahamian aragonite (7) which was very similar to Fredonia in particle size and SO_2 removal. Natural impurities associated with the aragonite caused severe foaming problems that would have to be solved before it could be considered for FGD. An unusual aspect of those tests was the rapid settling and high filterability of the slurry which suggested that the sludge quality may also be associated with limestone type. These slurry properties were, therefore, included in the special type-and-grind tests reported here.

Although the effects of type-and-grind have been clearly established for the venturi scrubber, the effects are less certain in higher performance systems such as a TCA. It would be expected that the effects of these variables would be increased by some degree as a result of better mass transfer on the liquid side. The single-loop turbulent contact absorber (TCA) used here is more representative of conventional FGD systems now in operation. These tests were made at RTP in preparation for more extensive evaluation to begin next year at the Shawnee Test Facility.

PROCEDURE

The principal objective of the tests was to compare the SO_2 removal of the limestone scrubber at standardized operating conditions to determine (a) the amount of coarse-ground limestone required (in terms of stoichiometric ratio) to equal the same SO_2 removal obtained with the same type of finely ground stone, and (b) whether the different types of limestones showed differences in scrubber performance when compared at equivalent grinds and stoichiometric ratios.

Limestone was ground by Allis-Chalmers Co., Milwaukee, WI. The pulverized-and-screened test stones were shipped to RTP in drums. The chemical and physical properties of the stones were determined in detail by Martin Marietta Laboratories, Baltimore, MD. Particle size distributions were checked by wet screening at RTP. The principal properties determined by these examinations are summarized in Tables 1 and 2 and elsewhere (8). Although three grinds of Fredonia were prepared, the "coarse" and "very coarse" actually showed little difference in particle size distribution. The other two stones were tested in fine and coarse grinds only.

A single-loop limestone scrubber was set up to provide test conditions selected by Bechtel. Those standard conditions were:

- TCA scrubber with internals to provide 20-cm H_2O pressure drop,
- 8 percent solids in recirculating slurry,
- No fly ash,
- 3.8 m/sec superficial gas velocity,
- 6.7 $l./m^3$ liquid-to-gas ratio,
- 3000 ppm SO_2 inlet,
- 5000 ppm Cl in liquor,
- 9 min. EHT residence time,
- No forced oxidation, and
- No chemical additives.

The TCA tower contained three beds of nitrile foam spheres and was operated at a slurry recirculation rate of 64 liters/min. Limestone makeup slurry was fed at several different rates to vary the stoichiometric ratio. This ratio is defined, for the purpose of this study, as:

$$\text{Stoichiometric ratio} = \frac{\text{moles } CaCO_3 \text{ fed}}{\text{moles } SO_2 \text{ absorbed}}$$

The principal response measured for these tests was SO_2 removal. Both wet sampling and ultraviolet analyses were used to determine the SO_2 concentration in the flue gas entering and leaving the scrubber. As a further check on these measurements, composite samples of the scrubber slurry solids were collected while operating at each of the limestone feed rates. Chemical analysis of the solids provided sulfur/calcium ratios that could be compared, via material balances, with the SO_2 removals determined by gas analysis.

Other variables monitored during the tests were: the percent solids in the filter cake, the settling rate, and the settled density of the scrubber slurry.

The amount of limestone available for the tests limited the total scrubber operating period to about 180 hours with each grind. The range of stoichiometric ratios and the number of data points were therefore constrained to the minimum, as was the ability to ensure true steady-state operation at each level of stoichiometry.

RESULTS

Fredonia Limestone

The SO_2 removals obtained with three nominally different grinds of Fredonia limestone are shown in Figures 1 and 2. Comparing performance at an SO_2 removal of 82 percent, the TCA scrubber required a stoichiometric ratio of 1.13 with fine grind, 1.47 with coarse, and 2.2 with very coarse.

An anomaly of this comparison is the large apparent difference between the responses for "coarse" and "very coarse" grinds: the difference cannot be explained by the particle size distributions, which were very similar for the two grinds (as determined by Martin Marietta Corporation and verified at RTP, Table 2). If the difference is assumed to be due entirely to experimental error, and a single curve is fitted between these two responses to represent coarse Fredonia, that curve would indicate that a stoichiometric ratio of 2.0 (±0.35) is required to equal the 85 percent SO_2 removal obtained with fine Fredonia at a stoichiometric ratio of 1.25. In either case, the effect of grind was greater than that assumed by the TVA study (5).

Georgia Marble

Figure 3 shows the effect of grind on the performance of the TCA scrubber with Georgia marble. These SO_2 removals were calculated from material balances because of leakage problems with the flue gas sampling system, which resulted in unreliable data from gas analyses (one datum, 76 percent removal at 1.63 stoichiometric ratio, was confirmed by the gas analyses). The fine-grind data indicate a much lower reactivity than Fredonia, which yielded 89 percent removal at 1.63 stoichiometric ratio. The comparison of fine and coarse grinds in Figure 3 indicates that stoichiometric ratios of about 1.3 and 2.3 are required, respectively, to achieve 63 percent SO_2 removal with Georgia marble.

Further comparison between Georgia marble and Fredonia limestone is made in Table 3 for the coarse grinds. These data show a consistent superiority of the Fredonia stone by all bases of measurement. The flue gas analyses show a difference (i.e., ΔSO_2 removal) of about 10 percent between the two types of stones while the solids analyses indicate a difference of about 15 percent. The reactivity of Georgia marble was thus significantly lower than Fredonia for both coarse and fine grinds. Since Georgia marble has the highest chemical purity of the test stones, it can be concluded that its lower reactivity must be related to some physical attribute, or "type."

Stone Man Limestone

The performance of the RTP scrubber with fine Stone Man is shown in Figure 4. About 74 percent SO_2 removal was obtained at a stoichiometric ratio of 1.5--marginally better than the performance with fine Georgia marble (70 percent removal) but still significantly less than that obtained with Fredonia limestone (88 percent removal) at the same stoichiometric ratio.

Although there were insufficient data to make a direct comparison of the stoichiometric ratio of fine Stone Man required to equal the SO_2 removal obtained with fine Fredonia, it is obvious that the difference between them is considerably greater than that observed at Shawnee (1.2 for Fredonia vs 1.3 for Stone Man). Since fine Fredonia yielded 84 percent removal at RTP with a stoichiometric ratio of 1.2, it is clear from

Figure 4 that Stone Man must require a stoichiometric ratio greater than 1.5 to achieve the same SO_2 removal. Figure 5 attempts a more definitive comparison of the three types of stones at equal SO_2 removal (of 84 percent). It extrapolates the data for Georgia marble and Stone Man (assuming the same slope as Georgia marble) to provide estimates of 1.86 and 1.76, respectively, as the stoichiometric ratios required to equal the SO_2 removal obtained with Fredonia limestone at a stoichiometric ratio of 1.2.

Coarse Stone Man gave the results shown in Figure 6. The effect of grind on SO_2 removal was much less than that observed with the other stones; at a stoichiometric ratio of 1.5, about 70 percent SO_2 removal was obtained with coarse Stone Man vs 74 percent with the fine grind. The results nevertheless indicate a significant difference in stoichiometric ratios required by coarse and fine grinds for equal SO_2 removal: at 74 percent removal, the stoichiometric ratios are 1.7 and 1.5, respectively.

Figure 6 also compares the performance of coarse Stone Man with the other two types of coarse limestones. The relative ranking of the coarse grinds was the same as that found for the fine grinds: Fredonia > Stone Man > Georgia marble. Unlike the comparison of fine grinds, however, the coarse Stone Man appears to resemble the performance of coarse Fredonia more than Georgia marble.

Slurry Properties

An interesting result of the tests with Georgia marble was the unusual properties of the scrubbing slurry. The most noticeable was the settling rate, which averaged 2.2 cm/min when feeding coarse grind and 2.6 cm/min when feeding fine-grind Georgia marble. This settling rate is 5 times faster than that obtained with Fredonia limestone. The settled density (0.46 g/ml) was also greater than that of the Fredonia tests (0.32 g/ml) and the filterability higher, yielding a filter cake containing 62 percent solids vs 54 to 58 percent solids when using Fredonia.

The unusual slurry properties obtained with Georgia marble were similar to those reported in previous RTP tests with aragonite (7), which settled at 3 cm/min to a density of 0.56 g/ml and filtered to 69 percent solids. These results--especially the rapid settling rates with both coarse and fine limestone grinds--show that the crystalline properties of the calcium sulfite precipitated in the limestone scrubber is affected by the type of limestone feed. This conclusion is supported by the absence of correlation between the stoichiometric ratio and settling rate for Georgia marble and Stone Man. The amount of unreacted limestone in the slurry, and its particle size, had little effect compared to the differences observed between settling rates of slurries obtained with the various types of limestones fed.

The properties of the scrubber slurries obtained with Stone Man limestone fell between those of the other two stones. The Stone Man slurries filtered to an average 59 percent solids and settled at

1.9 cm/min to a density of 0.46 g/ml. The ranking of stones with respect to sludge quality was thus the same as for SO_2 removal, but in reverse order; i.e., the lowest reactivity yielded the best sludge. The Stone Man slurries, like those obtained with Fredonia, also tended toward higher filterability and faster settling when using the coarse limestone grind.

DISCUSSION

The results of these tests agree with prior investigations that indicated that limestone type is an important factor in FGD scrubber performance. The lower SO_2 removals obtained with Stone Man, relative to Fredonia, can be only partially explained by its 10 percent lower content of available $CaCO_3$*. In the case of Georgia marble, the difference of type provides the only explanation for its lower performance.

The relative ranking of the three types was the same for coarse grinds as for fine grinds. This may not be generally true, since Shah (1) reported that relative rankings change with grind; two of his stones that were least reactive when ground to 325 mesh were most reactive when ground to 200 mesh. This strange effect tends to be supported by the two anomalies pointed out in the results discussed here: (a) the large apparent difference in SO_2 removals between "coarse" and "very coarse" Fredonia, in spite of the negligible difference in particle size distribution, and (b) the relatively small effect of particle size on the SO_2 removals obtained with Stone Man.

---

*For calculating stoichiometric ratios and limestone utilization, all of the $CaCO_3$ in the limestones is assumed to be available for reaction with SO_2.

There is no rational explanation for such an interaction between type and grind--one would expect the relative change in limestone performances to be equally affected by a given change in grind, and such an assumption is implicit when making economic comparisons of the effect of grind. As far as the RTP results are concerned, the data are not sufficiently extensive or consistent to accept (as fact) the two anomalies indicated above. The principal difficulty with the RTP test system, other than the measurement of SO_2 removal*, is ensuring the proper control of scrubber variables that affect SO_2 removal independently of limestone type, grind, and stoichiometry. Such variables include Cl^- concentration, which varied from 2000 to 5500 ppm in the scrubbing liquor; oxidation, which caused the sulfate relative saturation to vary from 10 to 60 percent; inlet SO_2 concentration, which varied from 2200 to 3300 ppm; and the L/G ratio. The last of these is probably the greatest unknown. Although the slurry recirculation rate is measured and (supposedly) readjusted each hour, changes in tower ΔP were observed to vary as much as 3 cm H_2O, indicating significant variation in the hydraulics--and therefore in SO_2 removal efficiency--of the scrubber.

---

*Measurement of SO_2 removal, as shown by the three different methods of SO_2 measurement used for these tests, can be deceptively difficult. The UV analyzer was the least reliable of the three. If material balances had not been used to verify flue gas analyses, completely different conclusions would have been drawn from the tests.

More accurate comparisons of limestone performance will require better control over all of these variables than has so far been achieved at RTP. Longer test periods of about 2 weeks at each stoichiometric ratio should also be provided to ensure that sufficient data are collected at truly steady-state conditions.

For the purpose of this study it therefore appears prudent to assume that the disagreement noted with regard to the effect of grind on SO_2 removal among the three stone types is due to experimental error rather than being an indication of intrinsic differences between the stones. In accordance with this assumption, the overall effect of a change in grind from 84 percent - 325 mesh to 70 percent - 200 mesh for the three stones required a change of stoichiometric ratio of about 0.6 to maintain a given SO_2 removal in the TCA scrubber. This compares to a value of 0.4 assumed in the TVA economic study (for a somewhat larger change in particle size) and therefore supports the major premise of the TVA study showing the economic benefit of fine grinding. In terms of limestone feed rate, the pilot plant tests indicate that the coarse grinds require about 50 percent more limestone to attain the same SO_2 removal as the fine grinds.

The SO_2 removal in the TCA scrubber was more sensitive to limestone stoichiometry than was the case with Shah's two-stage venturi. The six plots of SO_2 removal vs stoichiometric ratio made in this study had an average slope of 29 percent SO_2 removal per unit change in stoichiometry (standard deviation = 12). This compares with an average slope of 9 for

Shah's data. The data reported by Gleason (9) for a venturi/packed tower show a higher sensitivity than the TCA: 60 percent SO_2 removal per unit change in stoichiometry. The Shawnee two-loop venturi/spray tower data* are in general agreement with Gleason's value (cf. Runs 809-1A through 814-1A). The increasing sensitivity to limestone stoichiometry presumably reflects the decreasing importance of SO_2 absorption and the increasing importance of limestone dissolution as the dominant resistances when going from a venturi to a TCA and then to a two-loop system. One could thus speculate that the use of additives such as adipic acid could significantly reduce the adverse effects of poor quality limestones in conventional scrubbers.

The larger apparent effect of limestone type in the RTP scrubber, when comparing Fredonia and Stone Man, may be partly due to better mass transfer characteristics relative to the Shawnee two-stage system that was used in Bechtel's comparison (6). The spray tower was operated at relatively low L/G during those tests, so that the system may have been mass-transfer limited, in which case limestone type-and-grind effects would probably be less noticeable.

Relative Saturation

Using the Bechtel/Radian equilibrium program, $CaSO_4 \cdot 2H_2O$ saturations were computed from the analytical compositions of the scrubber feed liquor. All values were less than 1.0, indicating that the natural tendency of the TCA scrubber was to operate in the unsaturated mode regardless of the type-and-grind of the limestone feed. The O_2 content of the flue gas averaged 6.6 percent throughout the 4-month test period. The overall-average saturation for the coarse grinds was 39 percent with 18 percent oxidation while the fine grinds averaged 28 percent saturation with 12 percent oxidation. Fredonia limestone tended to show higher values of both variables while Stone Man showed the lowest values, but the data are not sufficiently extensive or consistent to establish any effect of limestone type.

---

*at a total L/G of 13 $l./m^3$

CONCLUSIONS

- The RTP results confirm that both type and grind of the limestone feed can significantly affect SO_2 removals obtained in a high-performance scrubber (TCA) of the single-loop type.

- The observed effects of limestone grind support the premise of TVA's economic study which showed clear benefits for fine grinding.

- The relative performance of the three stones tested was: Fredonia > Stone Man > Georgia marble. This ranking was valid for both fine and coarse grinds.

- For the three stones tested, sludge quality was also related to the type of limestone used. The settling rate and filterability of the scrubber slurry increased as SO_2 reactivity of the limestone decreased.

- With 6.6 percent O_2 and 3000 ppm SO_2 in the flue gas, the propensity of the scrubber to operate in the unsaturated mode was unaffected by limestone type or grind.

REFERENCES

1. Shah, I.S., "SO_2 Removal Using Calcium Based Alkalies: Pilot Plant Experience," Presented at 2nd International Lime/Limestone Wet Scrubbing Symposium, New Orleans, LA, November 1971.

2. Drehmel, D.C., "Limestone Types for Flue Gas Scrubbing," Presented at 2nd International Lime/Limestone Wet Scrubbing Symposium, New Orleans, LA, November 1971.

3. Burbank, D.A., and S.C. Wang, "EPA Alkali Scrubbing Test Facility: Advanced Program," Draft final report for EPA Contract 68-02-1814, Bechtel National, Inc., September 1979.

4. Burbank, D.A., et al., "EPA Alkali Scrubbing Test Facility, TVA Shawnee Power Plant, Paducah, KY," Progress report for period March 1 - March 31, 1979 by Bechtel National, Inc. for Contract 68-02-3114, pp. 1-2 to 1-9, May 9, 1979.

5. Little, A.F., "Preliminary Evaluation of the Effect of Limestone Grind Upon the Economics of Limestone Scrubbing FGD Facilities," TVA letter report of March 6, 1977 to J.E. Williams, EPA/IERL-RTP.

6. Head, H.N., et al., "EPA Alkali Scrubbing Test Facility, TVA Shawnee Power Plant, Paducah, KY," Progress report for period May 1 - May 31, 1979 by Bechtel National, Inc. for Contract 68-02-3114, p. 2-5, July 2, 1979.

7. Borgwardt, R.H., et al., "Limestone Scrubbing of SO_2 at EPA/RTP Pilot Plant," Progress Report 22, January 1976.

8. Borgwardt, R.H., et al., "Limestone Scrubbing of SO_2 at EPA/RTP Pilot Plant," Progress Reports 39, 40, 41, and 42, March-November 1979.

9. Gleason, R.J., "Limestone Scrubbing Efficiency of Sulfur Dioxide in a Wetted Film Packed Tower in Series with a Venturi Scrubber," Presented at 2nd International Lime/Limestone Wet Scrubbing Symposium, New Orleans, LA, November 1971.

TABLE 1. COMPOSITION OF TEST STONES

Limestone Type	Weight Percent (dry)			
	Total $CaCO_3$	Available[a] $CaCO_3$	$MgCO_3$	Insoluble Inerts
Georgia Marble	97.8	96.5	1.1	1.1
Fredonia	96.7	95.2	1.3	2.0
Stone Man	91.2	85.7	4.6	4.2

---

[a] assuming all magnesium to be present as dolomite, $MgCO_3 \cdot CaCO_3$

TABLE 2. PARTICLE SIZE OF TEST STONES

Limestone Type	Weight Percent		
	-325 mesh	-200 mesh	-100 mesh
Fredonia (fine)	84.1	95.3	97.7
Ga. Marble (fine)	84.1	98.6	99.8
Stone Man (fine)	82.9	98.7	98.8
Fredonia (coarse)	54.5	69.6	92.9
(very coarse)	51.2	65.0	88.4
Ga. Marble (coarse)	48.1	70.1	94.7
Stone Man (coarse)	52.9	71.6	96.4

TABLE 3. PERFORMANCE OF IERL-RTP SCRUBBER WITH COARSE-GRIND LIMESTONES

Limestone Type	Test Period, 1979	Limestone Feed Rate, kg/hr (dry)	SO_2 Feed Rate kg/hr	Stoichiometric Ratio[a]	SO_2 Removal, %		Limestone Utilization, %[b]
					UV	Wet	
Fredonia	July 10-12	8.63	3.08	2.16	85	80[c]	40.5
Ga. Marble	July 19-23	9.25	3.64	2.08	76	70[c]	38.5

[a] Based on UV analyses

[b] By solids analyses

[c] Best agreement with material balance

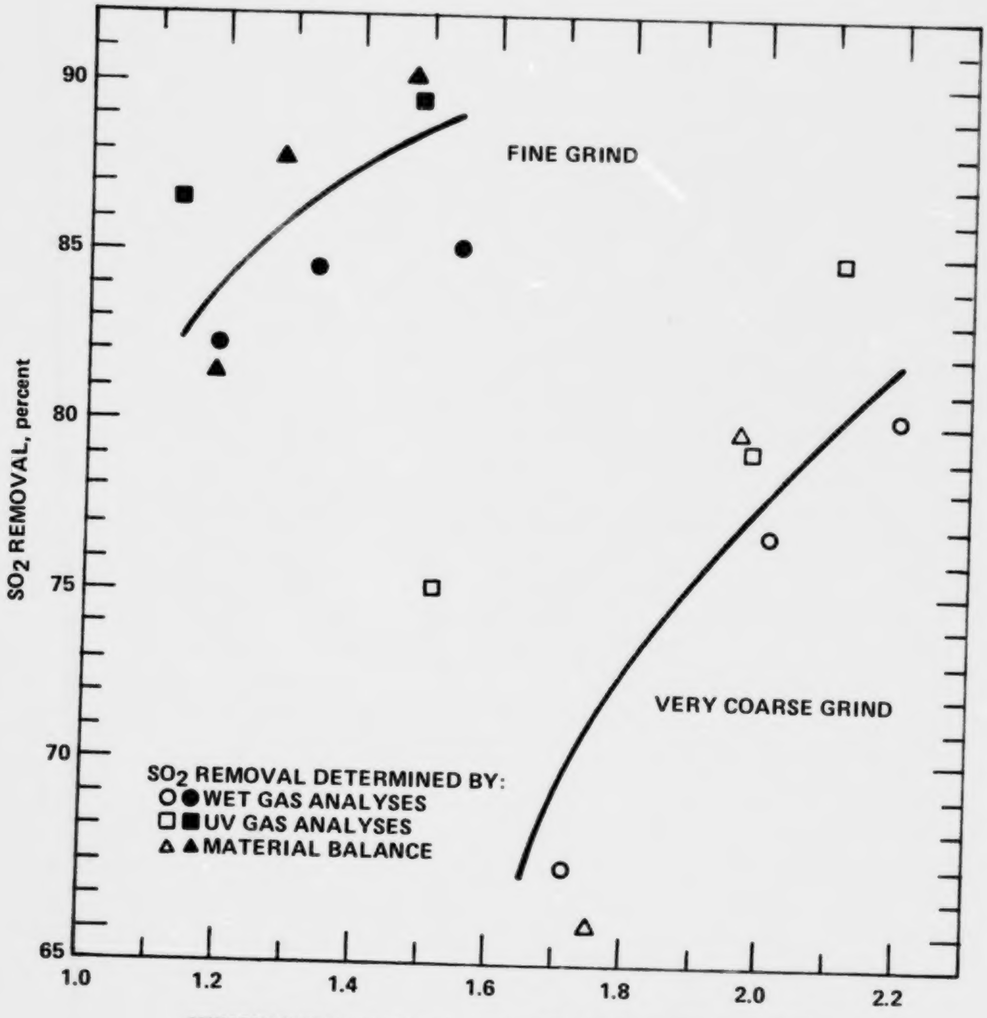


Figure 1. Performance of IERL-RTP pilot TCA scrubber with Fredonia limestone.

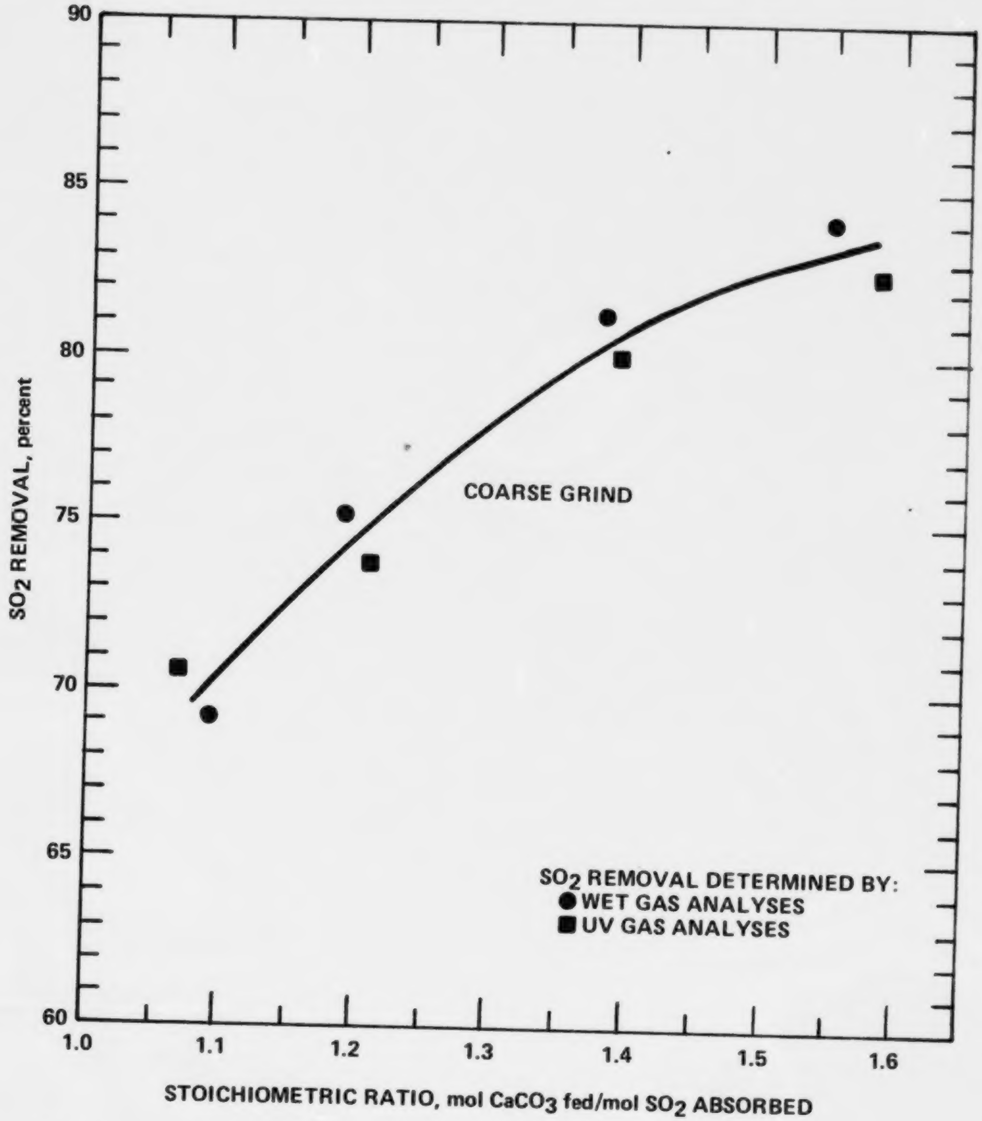


Figure 2. Performance of IERL-RTP pilot TCA scrubber with Fredonia limestone.

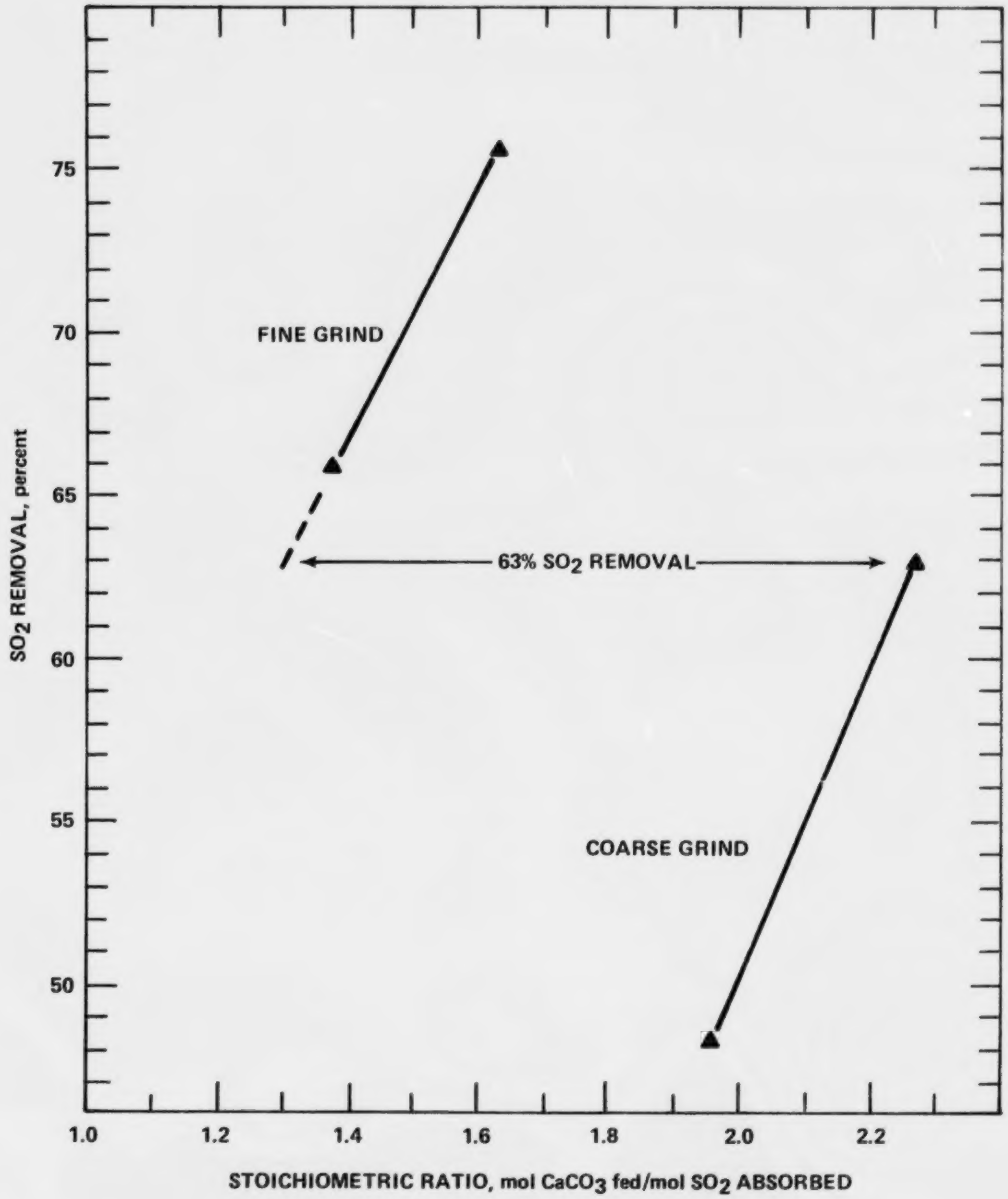


Figure 3. Performance of IERL-RTP pilot TCA scrubber with Georgia marble.

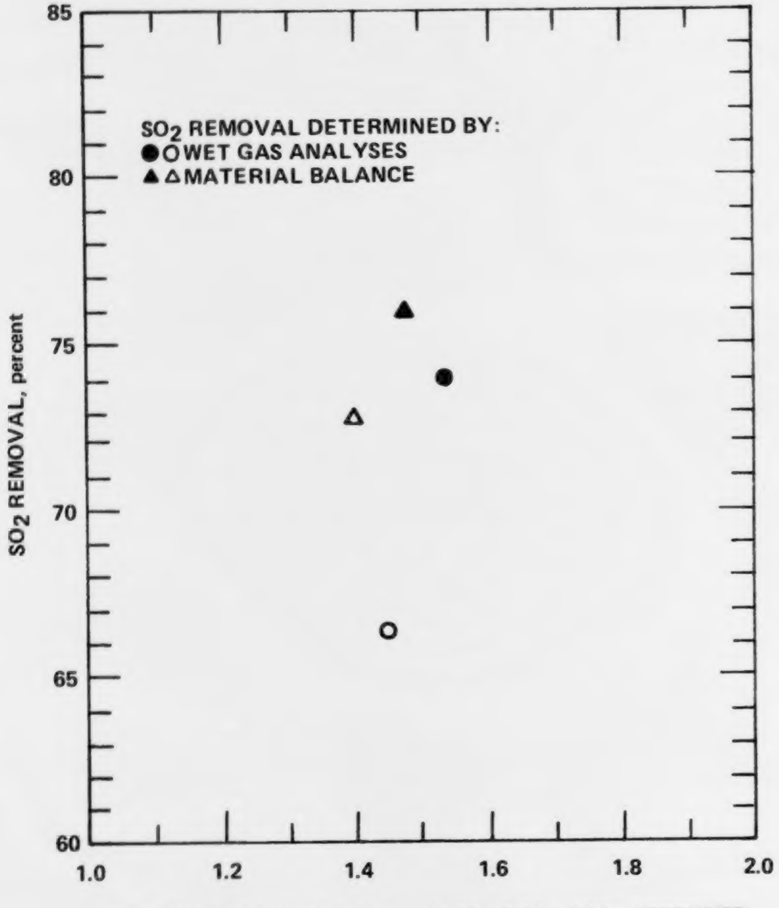


Figure 4. Performance of IERL-RTP pilot TCA scrubber with fine-grind Stone Man limestone.

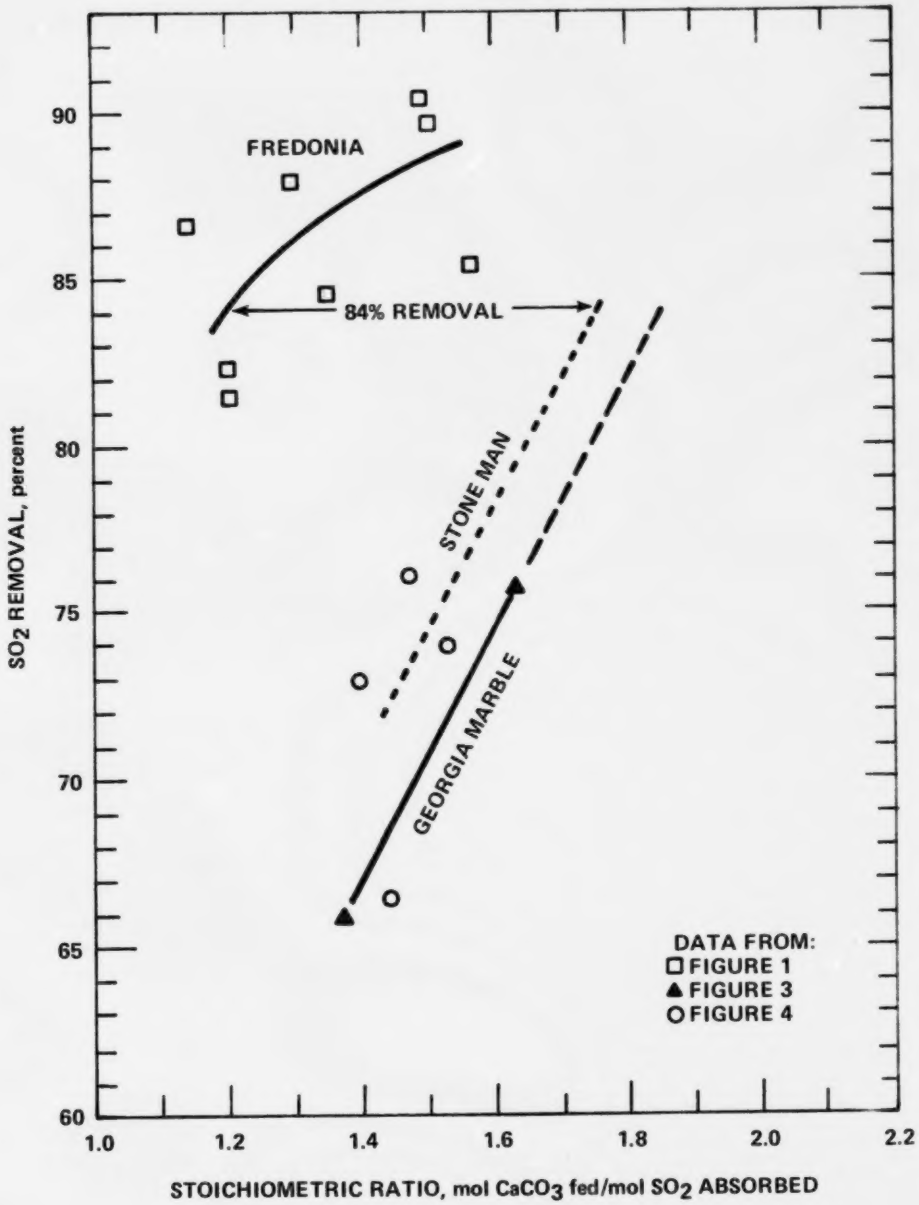


Figure 5. Comparison of IERL-RTP pilot TCA scrubber performances with three types of fine-grind limestones.

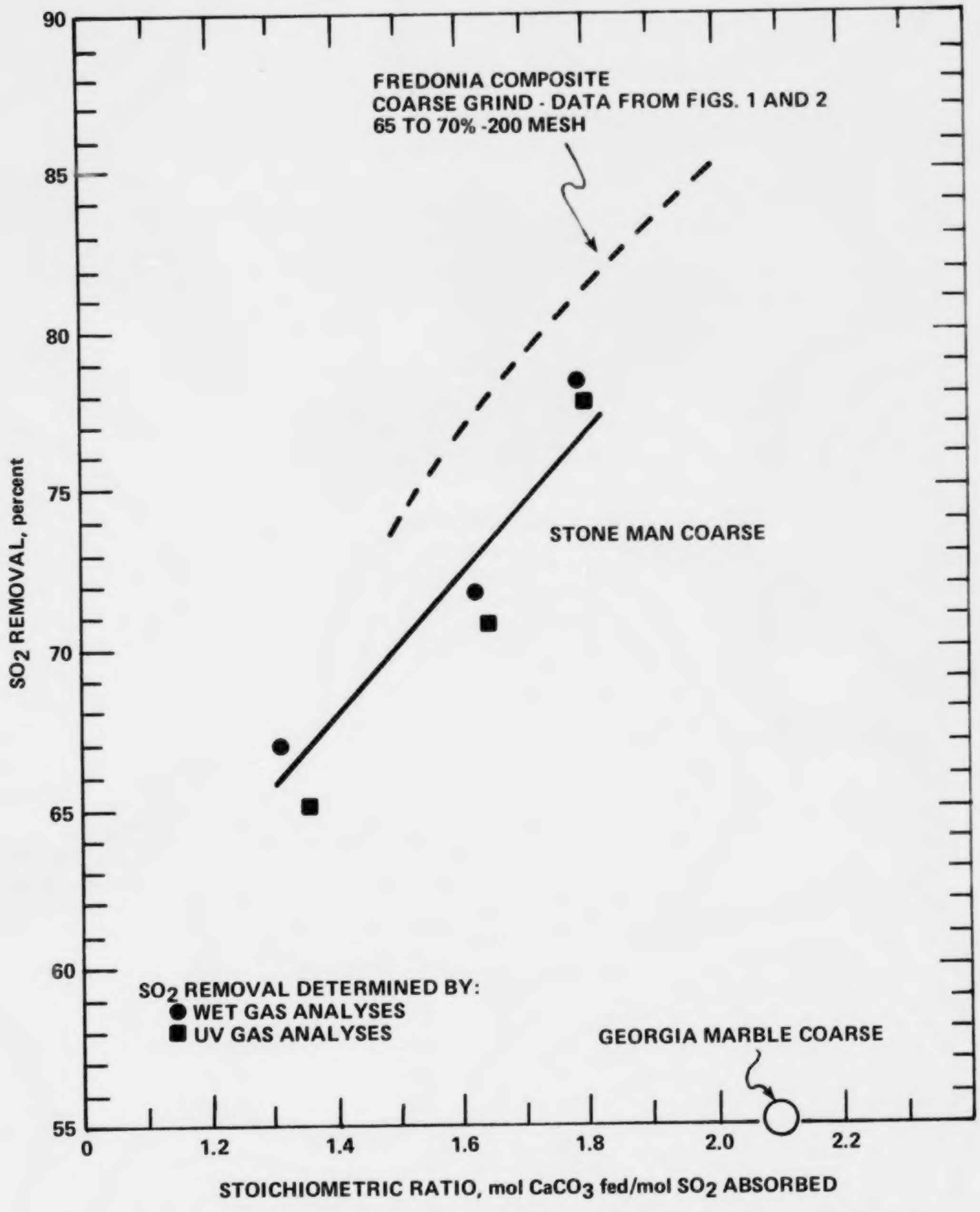


Figure 6. Comparison of IERL-RTP pilot TCA scrubber performances with three types of coarse-grind limestones.

TEST RESULTS ON ADIPIC ACID-ENHANCED LIME/LIMESTONE SCRUBBING AT THE EPA SHAWNEE TEST FACILITY

- SECOND REPORT -

by

D.A. Burbank and S.C. Wang

Bechtel National, Inc.
50 Beale Street
San Francisco, California 94105

For Presentation at the Industry
Briefing on EPA Lime/Limestone
Wet Scrubbing Test Program

Raleigh, North Carolina

December 5, 1979

ABSTRACT

Adipic acid has been demonstrated as an effective scrubbing additive for enhancing SO_2 removal in lime and limestone wet scrubbing tests both at EPA's IERL pilot plant at Research Triangle Park, North Carolina, and at the EPA-sponsored Shawnee Test Facility near Paducah, Kentucky.

Earlier test results using adipic acid, from July 1978 through late January 1979, have been reported in the Fifth Symposium on Flue Gas Desulfurization in Las Vegas, Nevada, March 5-8, 1979. This is a second report on the adipic acid test results at the Shawnee Test Facility from February through October 1979.

In the recent test period, wider ranges of adipic acid concentrations (0 to 4000 ppm) and other operating parameters have been explored, both with and without forced oxidation of calcium sulfite to calcium sulfate. Major efforts included conducting factorial tests to investigate the effects of adipic acid concentration and system operating parameters on SO_2 removal in both venturi/spray tower and TCA systems to quantify the scrubber performances. Preliminary equations for predicting SO_2 removal have been developed for the venturi scrubber operating with adipic acid-enhanced lime slurry and for the TCA with adipic acid-enhanced limestone slurry.

Other testing, without adipic acid additive, included the simulation of TVA's Widows Creek full-scale forced oxidation operation on the venturi/spray tower system, and the study of the effect of air stoichiometry and agitator speed on sulfite oxidation on the TCA system.

ACKNOWLEDGMENTS

The authors wish to express their appreciation to Dr. Gary Rochelle, consultant to the project, for his constant input of ideas and technical assistance in the adipic acid development program and to the following Bechtel personnel who contributed to the preparation of this paper:

C.L. DaMassa
J. Hing
T.M. Martin
R.R. McKinsey
D.T. Rabb
L.S. Reider
C.H. Rowland
M.A. Smith
V.C. Van der Mast
K.E. Wong

Further acknowledgment and appreciation are extended to the TVA staff at the Shawnee Test Facility and to TVA's Emission Control Development Projects group at Muscle Shoals, Alabama, who are responsible for operation, maintenance, and engineering modification of the facility.

NOTE

Although it is the policy of the EPA to use the metric system for quantitative descriptions, the British system is used in this report. Readers who are more accustomed to metric units are referred to the conversion table in the Appendix.

TABLE OF CONTENTS

Section		Page
1.	Introduction	34
2.	Bleed Stream Oxidation of Limestone/Adipic Acid Slurry from the Venturi/Spray Tower System	38
3.	Factorial Tests on the Venturi Scrubber with Lime/Adipic Acid Slurry	44
4.	Factorial Tests on the Venturi/Spray Tower System with Limestone/Adipic Acid Slurry	49
5.	Simulation of Widows Creek Full-Scale Forced Oxidation on the Venturi/Spray Tower System	57
6.	Lime/Adipic Acid Testing on the TCA with One Tank and No Forced Oxidation	62
7.	Lime/Adipic Acid Testing on the TCA with One Tank and Forced Oxidation	65
8.	Limestone/Adipic Acid Testing on the TCA with One Tank and Forced Oxidation	71
9.	Limestone/Adipic Acid Testing on the TCA with Two Tanks and Forced Oxidation	76
10.	Factorial Tests on the TCA System with Limestone/Adipic Acid Slurry	82
11.	Other Testing on the TCA System without Adipic Acid Additive	94
12.	Dewatering Characteristics of Adipic Acid-Enhanced Lime/Limestone Slurry	103
13.	Summary of Characteristics of Adipic Acid as a Scrubber Additive	106
14.	Future EPA Test Program for 1980	109
15.	References	111
	APPENDIX - Converting Units of Measure	113

ILLUSTRATIONS

Figure		Page
1-1	Schedule for Advanced Test Program - 1979	37
2-1	Flow Diagram for Bleed Stream Oxidation in the Venturi/ Spray Tower System	39
2-2	Arrangement of Venturi/Spray Tower Oxidation Tank	40
3-1	Effect of Adipic Acid Concentration on SO_2 Removal in the Venturi with Lime Scrubbing	46
3-2	Percent SO_2 Removal vs. Adipic Acid Concentration During Adipic Acid Depletion Lime Runs VAA112 and VAA114	48
4-1	Flow Diagram for Adipic Acid Enhanced Scrubbing in the Venturi/Spray Tower System with Two Scrubber Loops and Forced Oxidation	50
4-2	Effect of Adipic Acid Concentration and Spray Tower L/G on SO_2 Removal in the Spray Tower	54
4-3	Effect of Adipic Acid Concentration and L/G in Spray Tower on SO_2 Removal in the Venturi/Spray Tower	56
5-1	Flow Configuration for Widows Creek Simulation on the Venturi/Spray Tower with Backmix	58
7-1	Flow Diagram for Adipic Acid-Enhanced Lime or Limestone Scrubbing in the TCA System with One Tank and Forced Oxidation	66
7-2	Arrangement of the TCA Oxidation Tank	67
9-1	Flow Diagram for Adipic Acid-Enhanced Limestone Scrubbing in the TCA System with Two Tanks and Forced Oxidation	77
10-1	Measured SO_2 Removal vs. Predicted SO_2 Removal by Equation 10-1	87
10-2	Effect of Adipic Acid Concentration and Slurry Flow Rate on SO_2 Removal in the TCA with Four Grids and Without Spheres	88
10-3	Effect of Adipic Acid Concentration and Slurry Flow Rate on SO_2 Removal in the TCA with Four Grids and 15 inches of Spheres	89

Figure		Page
10-4	Effect of Adipic Acid Concentration and Scrubber Inlet pH on SO_2 Removal in the TCA with Four Grids and 15 Inches of Spheres	91
10-5	Effect of Scrubber Inlet pH and Adipic Acid Concentration on SO_2 Removal	93
11-1	Effect of Air Stoichiometry and Agitator Speed on Sulfite Oxidation	100

TABLES

Table		Page
2-1	Results of Venturi/Spray Tower Limestone/Adipic Acid Tests with Bleed Stream Oxidation	41
3-1	Results of Lime/Adipic Acid Factorial Tests Using the Venturi Only	45
4-1	Results of Limestone/Adipic Acid Factorial Testing on the Venturi/Spray Tower System Using High Sulfur Coal	51
4-2	Results of Limestone/Adipic Acid Factorial Testing on the Venturi/Spray Tower System Using Low Sulfur Coal	52
5-1	Results of Venturi/Spray Tower Limestone Tests with Two-Scrubber-Loop Forced Oxidation and Backmix - Widows Creek Simulation	59
6-1	Results of TCA Lime/Adipic Acid Tests Without Forced Oxidation	63
7-1	Results of TCA Lime/Adipic Acid Tests with Forced Oxidation	68
8-1	Results of TCA Limestone/Adipic Acid Tests with Forced Oxidation Using One Tank	72
9-1	Results of TCA Limestone/Adipic Acid Tests with Forced Oxidation Using Two Tanks	77
10-1	TCA Limestone/Adipic Acid Factorial Test Results	83
11-1	TCA Lime Long-Term Reliability Test	95
11-2	Results of TCA Limestone Tests with Forced Oxidation Using One Tank - Effect of Air Stoichiometry and Agitator Speed	97
11-3	Results of Limestone Factorial Tests with Forced Oxidation Using One Tank - Low Inlet SO_2 Concentration	102
12-1	Comparison of Shawnee Waste Slurry Dewatering Characteristics With and Without Adipic Acid Addition	104

Section 1

INTRODUCTION

One of the primary objectives of the Environmental Protection Agency (EPA) alkali wet scrubbing test program during the last 17 months has been to enhance the SO_2 removal and improve the reliability and economics of lime and limestone wet scrubbing systems by use of adipic acid as a chemical additive.

This is the second report addressing the results of testing with adipic acid from February through October 1979 at EPA's prototype Shawnee Test Facility located at the Tennessee Valley Authority (TVA) Shawnee Power Station near Paducah, Kentucky. Previous test results with adipic acid from July 1978 through late January 1979 at the Shawnee Test Facility and at EPA's pilot plant at Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina (IERL-RTP), have been reported in References 1 and 2.

Adipic acid is a dicarboxylic organic acid [$HOOC(CH_2)_4COOH$], which is commercially available and used mainly as a raw material in the nylon manufacturing industry. It effectively buffers the pH* in limestone/lime SO_2 absorbers and improves the SO_2 removal efficiency. The theoretical basis for its effect on the performance of limestone and lime scrubbers was developed in detail by Rochelle (Reference 3): the buffering action limits the drop in pH that normally occurs at the gas/liquid interface during SO_2 absorption, and the higher concentration of SO_2 in the surface film resulting from this buffering action accelerates the liquid-phase mass transfer. The capacity of the bulk liquor for reaction with SO_2 is also increased by the presence of calcium adipate in solution. SO_2 absorption is therefore less dependent upon the dissolution of limestone or $CaSO_3$ in the absorber. The overall result of these effects is improved SO_2 removal in limestone or lime scrubbers of a given type operating at a given liquid-to-gas ratio (L/G). In the case of limestone, it follows that a given SO_2 removal efficiency can be achieved at a lower stoichiometric ratio.

---

*Adipic acid has two buffering points. In the absence of chloride these are approximately pH 5.5 and 4.5. Chloride concentrations in the range of 5000 to 7000 ppm depress these buffering points to about pH 5.0 and 4.0.

Further analysis by Rochelle (Reference 4) indicated that the use of additives would be most attractive economically when used in scrubbers employing forced oxidation. If no decomposition or volatilization of the additive occurs, the makeup requirements for the additive would be minimized by the tightly closed loop resulting from the better dewatering properties of oxidized sludge. For this reason, the scrubber tests reported here emphasize the evaluation of adipic acid in combination with forced oxidation.

Adipic acid has several advantages over other additives, such as MgO, which are also known to increase SO_2 removal, but do so by means of reactions involving the sulfite/bisulfite equilibrium. Since adipic acid does not depend on this mechanism for its buffering activity, it is not adversely affected by oxidation of the sulfite in the scrubbing liquor. It should therefore be especially useful in single loop scrubbers that employ forced oxidation. Furthermore, the buffering mechanism by which adipic acid enhances SO_2 absorption is not affected by the presence of chloride. The chloride depresses the pH buffer points of adipic acid as previously mentioned, but the rate of drop in pH buffer points is non-linear and decreases with increase in chloride concentration. The lack of interference by chloride means that adipic acid should be fully effective in the most tightly closed loop systems.

The optimum concentration of adipic acid for effective improvement in SO_2 removal is only 5 to 10 millimoles/liter (700 to 1500 ppm). Even with the degradation of the acid, which results in an actual addition rate 3 to 5 times the theoretical requirement as experienced thus far, the preliminary economic evaluations (Reference 1) have shown that adipic acid can reduce the operating cost of limestone systems while simultaneously increasing performance.

A more detailed economic study, comparing the costs of limestone and lime FGD systems with respect to adipic acid addition, MgO addition, and forced oxidation, has been initiated by EPA. The study will be conducted by TVA with assistance from Bechtel. This study will be described in more detail by TVA later in this Briefing.

Testing with adipic acid at the Shawnee Test Facility was initiated in July 1978 and has continued since as a major part of the Shawnee Advanced Test Program. The results of testing with adipic acid additive since February 1979 are presented in the following sections, along with other test results without adipic acid such as the simulation of TVA's Widow Creek full-scale oxidation operation on the venturi/spray tower system.

System configurations successfully demonstrated during this period are:

- Adipic acid-enhanced limestone or lime scrubbing in a venturi/spray tower system with two scrubbing loops and forced oxidation in the first loop.

- Bleed stream oxidation of adipic acid-enhanced limestone slurry from venturi/spray tower system.

- Adipic acid-enhanced limestone or lime scrubbing in a Turbulent Contact Absorber (TCA) system with or without forced oxidation.

Other configurations have been used in the venturi/spray tower system (venturi or spray tower only operation) during factorial tests. However, because of the nature of the testing, the ranges of operating conditions selected were sometimes extreme by necessity and were not always conducive to system reliability.

TEST FACILITY AND PROGRAM

There are two scrubber systems operating at the EPA sponsored Shawnee Test Facility, each with its own independent slurry handling facilities. Both systems were tested with adipic acid additive. The systems have the following scrubbers:

- A venturi followed by a spray tower (V/ST)
  (35,000 acfm capacity @ 300°F)
- A Turbulent Contact Absorber (TCA)
  (30,000 acfm capacity @ 300°F)

The flue gas capacity of each scrubber system is equivalent to 10 MW of boiler capacity. The scrubbers receive flue gas from TVA Shawnee coal-fired boiler No. 10. The boiler normally burns a high-to-medium sulfur bituminous coal producing SO_2 concentrations of 1500 to 4500 ppm. Flue gas can be taken from either upstream or downstream of the boiler No. 10 particulate removal equipment, allowing testing with high fly ash loadings (3 to 6 grains/scf dry) or low loadings (0.04 to 0.2 grains/scf dry).

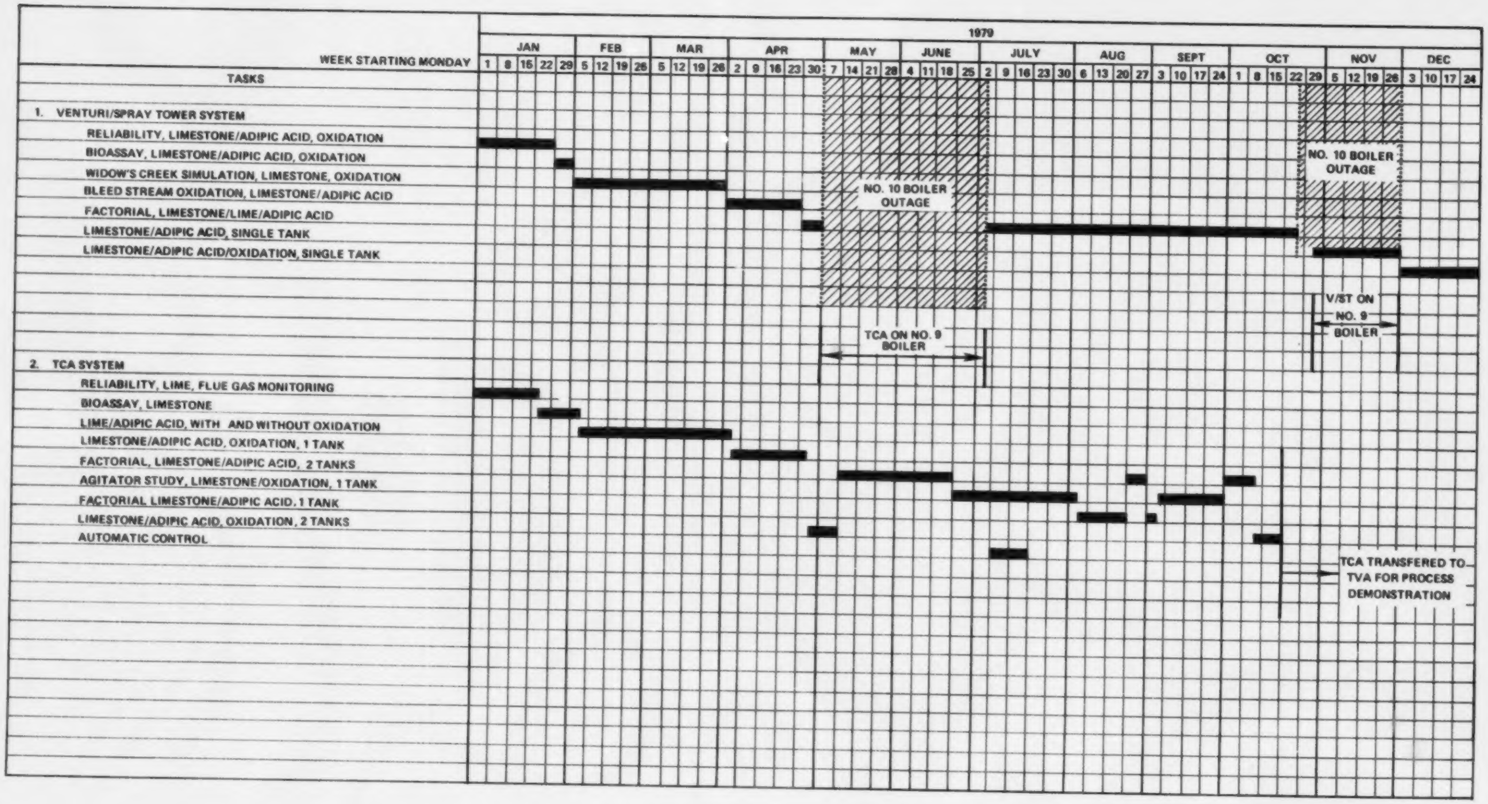
It is important to recognize that all tests in the adipic acid series were made with high fly ash loadings.

The test program was conducted with the scrubbing loop fully closed. Chlorides from the flue gas were concentrated in the scrubber slurry liquor over a range of 1000 to 10,000 ppm depending on the tightness of the scrubber water balance and the chloride concentration in the coal burned.

The Shawnee Test Facility has been operating since March 1972. Bechtel National, Inc. of San Francisco is the major contractor and test director; TVA is the constructor and test facility operator. The initial test program lasted through October 1974 (Reference 5) with the major emphasis on demonstrating reliable operation. The adipic acid and forced-oxidation tests are a part of an advanced test program that is currently scheduled to continue through December 1979. Earlier results of the advanced test program and a description of the test facility are reported elsewhere (References 6, 7, 8, 9, 10, 11 and 12).

The Advanced Test Program schedule for the period February through October 1979 covered in this report is shown in Figure 1-1. As can be seen, testing with adipic acid additive has constituted the major effort during this period.

Figure 1-1. SCHEDULE FOR ADVANCED TEST PROGRAM - 1979



Section 2

BLEED STREAM OXIDATION OF LIMESTONE/ADIPIC ACID SLURRY FROM THE VENTURI/SPRAY TOWER SYSTEM

One month of testing was conducted on the venturi/spray tower system in April 1979 (Figure 1-1) to investigate the feasibility of forced oxidation of calcium sulfite to gypsum in the system bleed slurry prior to dewatering. It was thought that the buffering capacity of adipic acid might maintain a sufficiently low pH in the oxidation tank to allow good oxidation. The bleed stream oxidation would be attractive for retrofit forced-oxidation installation where only an oxidation tank and the ancillary air sparging system and air compressor need be installed as add-on equipment.

SYSTEM DESCRIPTION

The flow diagram for the venturi/spray tower bleed stream oxidation is shown in Figure 2-1. The system was adapted from the two-scrubber-loop forced-oxidation configuration (Reference 1) by recirculating the venturi slurry to and from the same effluent hold tank for the spray tower. The system bleed stream was taken from the effluent hold tank and sent to the oxidation tank into which air was sparged through a simple 3-inch pipe ell. The arrangement of the oxidation tank is shown in Figure 2-2 which is drawn roughly to scale. The tank is 8 feet in diameter and was operated at the 18-foot level in all the bleed stream oxidation tests. The agitator in the tank has two downflow turbines. Each turbine is 52 inches in diameter and contains four blades. The bottom turbine is 10 inches above the air sparger. The agitator speed is 56 rpm and is rated at 17 brake Hp. Adipic acid was added to the effluent hold tank as a dry powder. The solids dewatering system in this series of tests consisted of a clarifier followed by a rotary drum vacuum filter.

DISCUSSION OF TEST RESULTS

Table 2-1 presents a summary of major test conditions and average results for the 5 runs made in this test series. Controlled run conditions common to all of the tests were:

Fly ash loading: High (3 to 6 grains/dry scf)
Venturi slurry rate: 600 gpm
Spray tower slurry rate: 1600 gpm

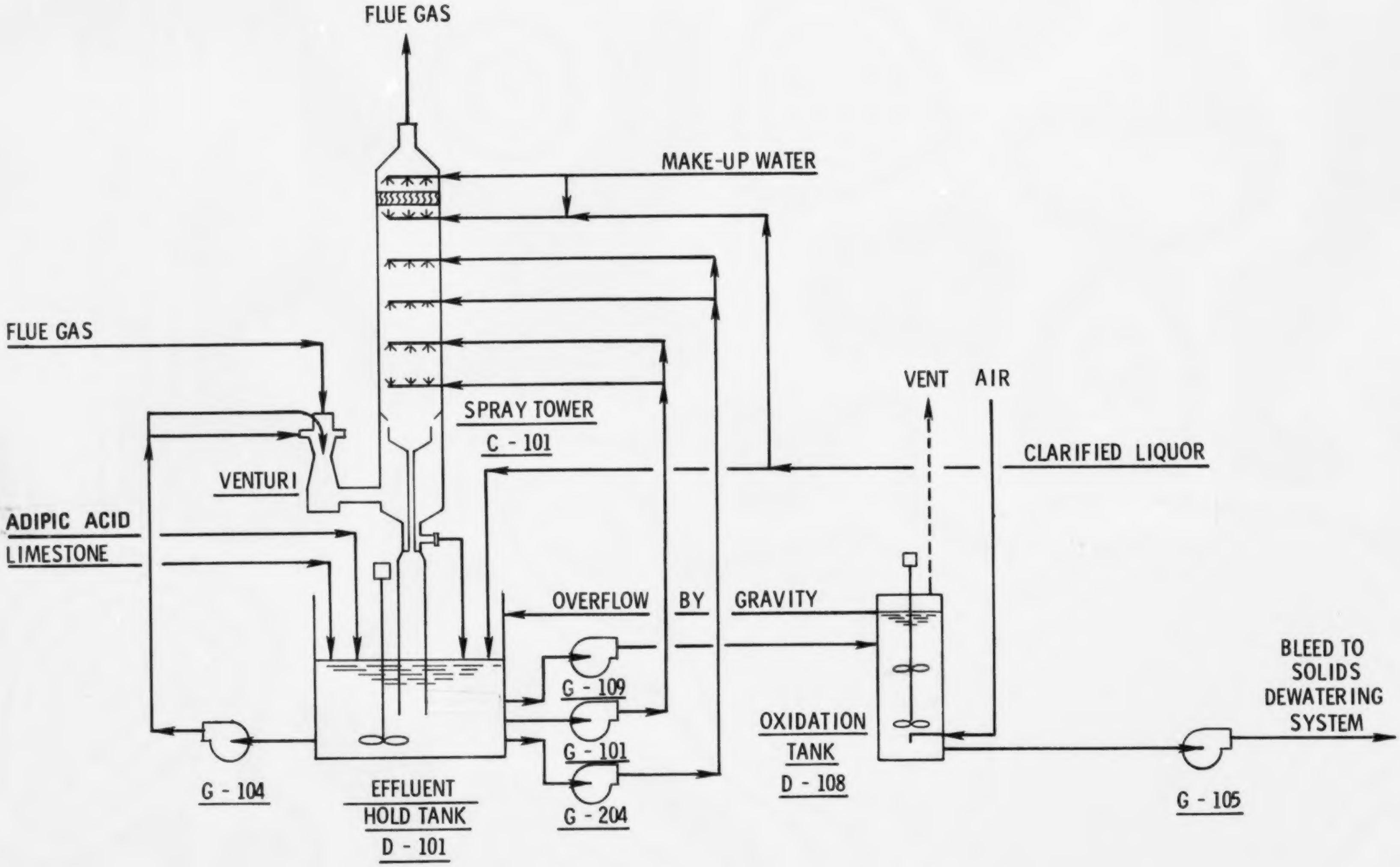
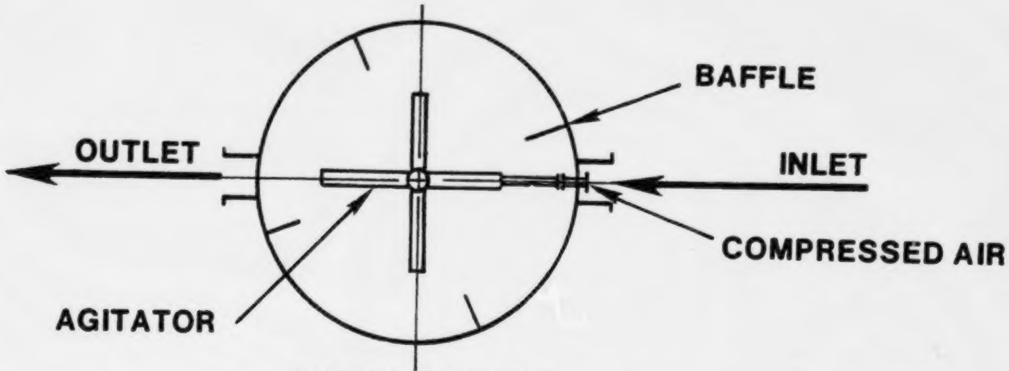
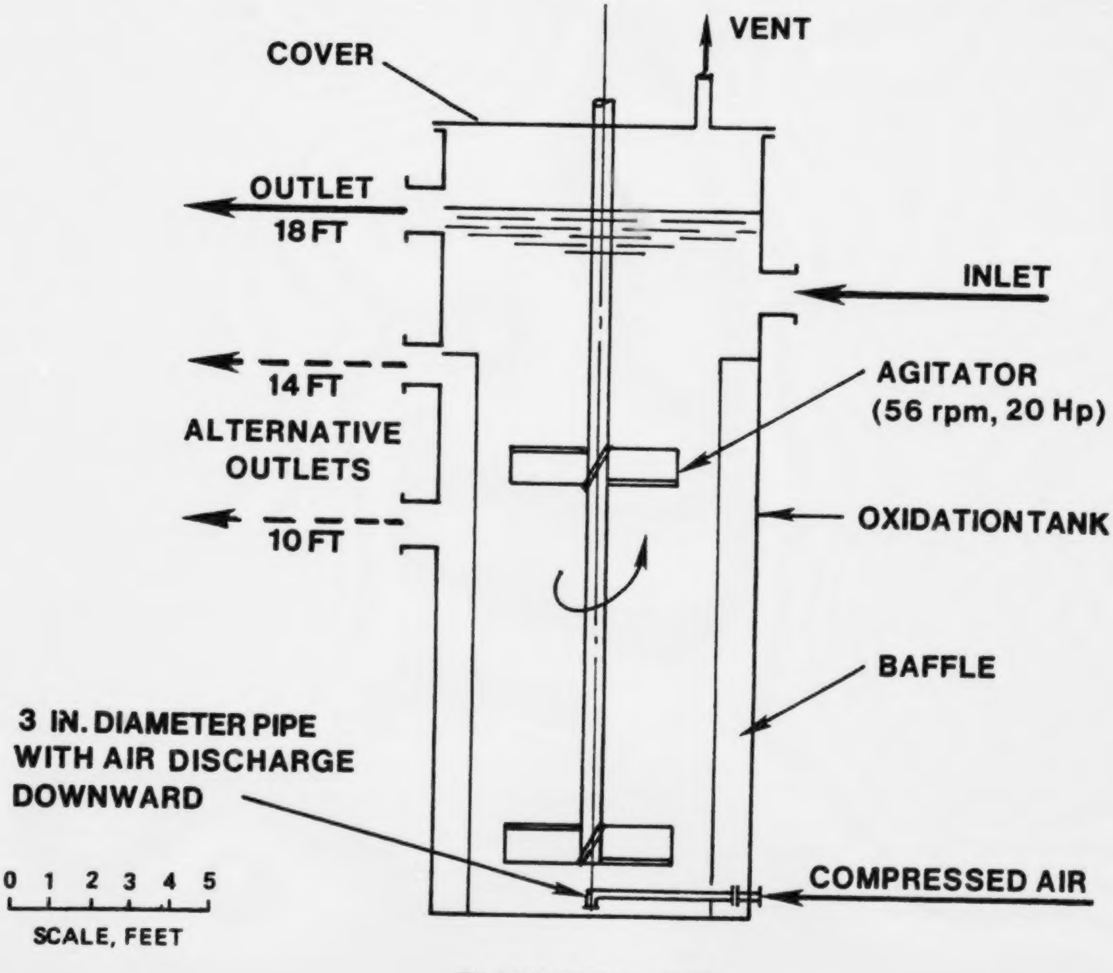

FLUE GAS
MAKE-UP WATER
FLUE GAS
VENT
AIR
SPRAY TOWER
C - 101
CLARIFIED LIQUOR
VENTURI
ADIPIC ACID
LIMESTONE
OVERFLOW BY GRAVITY
G - 109
G - 101
G - 204
G - 104
EFFLUENT
HOLD TANK
D - 101
OXIDATION
TANK
D - 108
BLEED TO
SOLIDS
DEWATERING
SYSTEM
G - 105

Figure 2-1. FLOW DIAGRAM FOR BLEED STREAM OXIDATION IN THE VENTURI/SPRAY TOWER SYSTEM


BAFFLE
OUTLET
INLET
COMPRESSED AIR
AGITATOR

OXIDATION TANK
PLAN VIEW


VENT
COVER
OUTLET
18 FT
INLET
14 FT
ALTERNATIVE
OUTLETS
10 FT
AGITATOR
(56 rpm, 20 Hp)
OXIDATION TANK
BAFFLE
3 IN. DIAMETER PIPE
WITH AIR DISCHARGE
DOWNWARD
COMPRESSED AIR
0 1 2 3 4 5
SCALE, FEET

ELEVATION VIEW

Figure 2-2. ARRANGEMENT OF VENTURI/SPRAY TOWER OXIDATION TANK (D-108)

Table 2-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE/ADIPIC ACID TESTS WITH BLEED STREAM OXIDATION

Major Test Conditions	909-1A	910-1A	911-1A	912-1A	913-1A
Fly ash loading	High	High	High	High	High
Gas rate, acfm @ 300°F	18,000	35,000	35,000	35,000	18,000
Venturi liquor rate, gpm	600	600	600	600	600
Spray tower liquor rate, gpm	1600	1600	1600	1600	1600
Percent solids recirculated (controlled)	15	15	15	15	15
Residence times (min)/Tank level (ft): EHT	11.8/11	11.8/11	11.8/11	11.8/11	11.8/11
Oxidation Tank	/18	/18	/18	/18	/18
Limestone stoich. ratio (controlled)	1.2	1.1	1.1	1.1	1.1(1)
Adipic acid concentration, ppm	1500	1500	1500	1500	1500
Air rate to oxidizer, scfm	120	240	240	350	120
Venturi pressure drop, Inches H_2O	9	9	7.5	7.5	8.0
Recycle flow from oxidation tk to EHT, gpm	0	0	60	0	0
Run-Average Results					
Start-of-run date	3/30/79	4/11/79	4/16/79	4/24/79	4/26/79
Onstream hours	239	127	134	53	23
Percent SO_2 removal	96	95	95	96	98
Inlet SO_2 concentration, ppm	2825	2550	2400	2650	2350
Adipic acid concentration, ppm	1475	1500	1575	1500	1625
Scrubber percent solids recirculated	12.1	15.3	15.0	14.0	13.5
Scrubber inlet pH	5.30	5.25	5.30	5.40	5.20
Oxidation tank pH	5.6	7.0	5.6	7.0	5.5
Limestone utilization	94	96	85	93	96
Sulfite oxidation in oxidation tk, %	95	71	99	40	95
Sulfite oxidation in scrubber inlet, %	37	28	70	33	36
Gypsum sat'n in oxidation tank, %	90	85	95	95	95
Gypsum sat'n in scrubber inlet, %	105	100	100	100	105
Oxidation tk liquor $SO_3^=$ concentration, ppm	70	95	40	125	125
Air stoichiometry, lb-atoms O/lbmole SO_2 absorbed	1.50	1.75	1.85	2.40	1.75
Filter cake solids, wt %	84	75	84	63	79
Mist eliminator restriction, %	-	2	-	-	0

Note: (1) To insure low pH & high oxidation the pH was controlled at 5.0 with 4.9 underride.

Slurry solids concentration: 15 percent
Effluent hold tank residence time: 11.8 min.
Oxidation tank level: 18 ft.
Adipic acid concentration: 1500 ppm
Nominal venturi pressure drop: 9 in. H_2O

The adipic acid concentration in the scrubber slurry loop was maintained at 1500 ppm* for all runs by adding adipic acid as a dry power in the effluent hold tank.

Run 909-1A. In the initial Run 909-1A, the flue gas rate was held at 18,000 acfm at 300°F, half the scrubber capacity. At this gas rate the bleed stream flow rate was about 15 gpm, compared to about 30 gpm normal bleed rate at full gas load. The oxidation tank residence time was therefore doubled to about 7.5 hours at 15 gpm, giving the bleed stream oxidation mode of operation the best chance of success. The results were encouraging. Although close attention was required to maintain the scrubber inlet pH at about 5.3, good control was achieved and oxidation of the bleed stream averaged 95 percent. Oxidation tank pH averaged 5.6. The air stoichiometry was 1.50 atoms oxygen/mole SO_2 absorbed. Good SO_2 removal of 96 percent was attained at an inlet SO_2 concentration in the flue gas of 2825 ppm. The adipic acid concentration was 1475 ppm.

Run 910-1A. In Run 910-1A, flue gas flow rate was increased to the full capacity of 35,000 acfm at 300°F. The corresponding bleed rate was about 30 gpm. The results showed that sufficiently low pH could not be held in the scrubber inlet slurry to give good oxidation without sacrificing SO_2 removal. Even during periods when the scrubber inlet pH averaged only 5.1, the oxidation tank pH was 6.9. For Run 910-1A, the average scrubber inlet pH (effluent hold tank pH) was 5.25 which gave an average SO_2 removal of 95 percent at an inlet SO_2 concentration of 2550 ppm. With the inlet pH to the oxidation tank at 5.25, the oxidation tank pH averaged about 7.0 and oxidation dropped off to 71 percent even though the air stoichiometric ratio was up to 1.75 atoms oxygen/mole SO_2 absorbed.

Run 911-1A. This run was made to see if a lower pH could be maintained in the oxidation tank by recycling 60 gpm of slurry back to the effluent hold tank. It was believed that this would force more of the low pH liquor in effluent hold tank through the oxidation tank. With a pH of 5.3 in the effluent hold tank, the oxidation tank pH was held at 5.6 and sulfite oxidation reached a high of 99 percent. Air stoichiometry was 1.85 atoms oxygen/mole SO_2 absorbed. With 1575 ppm adipic acid in the liquor, the SO_2

---

*Since some adipic acid decomposes to lower-carbon carboxylic acids and the analytical method employed at the Shawnee laboratory determines total carboxyl group, the "adipic acid concentration" reported throughout this report means "total carboxylic acid expressed as adipic acid."

removal averaged 95 percent at an inlet SO_2 concentration of 2400 ppm. Because of the 60 gpm recycle, limestone utilization was only 85 percent for Run 911-1A compared to 93 percent or higher for other runs. The recycle stream also resulted in 70 percent oxidation in the scrubber slurry compared to 37 percent or less for other runs.

Run 912-1A. In Run 912-1A, the recirculation of liquor from the oxidation tank was stopped and the air rate was increased to the maximum in an attempt to oxidize the bleed without artificially lowering the oxidation tank pH by the recycle stream. The increased air rate only compounded the problems experienced in Run 910-1A. Although the oxidation tank pH was again 7.0, on the average, the oxidation dropped to 40 percent. The air stoichiometry was 2.40 atoms oxygen/mole SO_2 absorbed. The SO_2 removal, however, remained high at 96 percent with 2650 ppm SO_2 in the flue gas.

Run 913-1A. The purpose of Run 913-1A was to operate the system under the conditions of Run 909-1A (18,000 acfm gas rate at 300°F) when good oxidation had been achieved, to allow Radian Corporation to take material balance samples around the oxidation tank. It is hoped that these samples will lead to additional knowledge of the decomposition mechanism of adipic acid in an oxidizing atmosphere. Good oxidation was again achieved with the lower bleed rate. At an average oxidation tank pH of 5.5, sulfite oxidation was 95 percent. The air stoichiometry was 1.75 atoms oxygen/mole SO_2 absorbed.

Although the mechanism by which adipic acid decomposes under scrubber environment is not completely understood, a preliminary study by Radian on samples taken during Run 913-1A showed further that a considerable amount of adipic acid and its degraded products (about 8 percent of feed) were "adsorbed" or "occluded" in the filter cake solids that could not be extracted by either strong acid or strong alkali solution (Reference 13).

SUMMARY

Bleed stream oxidation of limestone/adipic acid slurry is feasible if the oxidation tank pH can be maintained below about 6 with 1500 ppm of adipic acid acting as a pH buffer. The low pH can be accomplished by recycling the oxidation tank slurry (60 gpm) back to the effluent hold tank. It is also feasible without the recycle, but at a high oxidation tank residence time of about 7.5 hours for the bleed stream (at 1500 ppm adipic acid).

The recycle between the oxidation tank and the effluent hold tank essentially creates an operating mode approaching that of forced oxidation within the scrubber loop.

Section 3

FACTORIAL TESTS ON THE VENTURI SCRUBBER WITH LIME/ADIPIC ACID SLURRY

A series of 14 lime/adipic acid factorial tests (Runs VAA101 through VAA114) were conducted using the venturi scrubber only (spray tower slurry flow turned off). All tests were made without forced oxidation. The flue gas flow rate (liquid-to-gas ratio) and adipic acid concentration were varied to observe their effect on the SO_2 removal. Operating conditions held constant during these tests were:

Fly ash loading: High (3-6 grains/dry scf)
Venturi inlet pH: 6.0
Venturi pressure drop: 9 inches H_2O
Venturi slurry flow rate: 600 gpm
Slurry solids concentration: 8 percent
Hold tank residence time: 24 minutes

The test results are listed in Table 3-1.

The effect of adipic acid concentration on SO_2 removal has been correlated for 12 of the 14 runs (Runs VAA112 and VAA114 were adipic acid depletion tests) by the following equation:

$$\begin{array}{c}\text{Fraction } SO_2\\ \text{Removal}\end{array} = 1 - \exp\,[-0.48 \exp\,(3.3 \times 10^{-4} A)] \qquad (3\text{-}1)$$

where A is the adipic acid concentration in ppm.

Adipic acid concentrations were varied between 800 and 3200 ppm. Gas flow rate, which varied from 20,000 to 35,000 acfm (@ 300°F), had no significant effect on the SO_2 removal.

Equation 3-1 explains 89 percent of the variation in the data for SO_2 removal, with a standard error of estimate of 3.3 percent SO_2 removal.

Figure 3-1 shows SO_2 removal as a function of adipic acid concentration for the venturi with lime scrubbing. The curve in Figure 3-1 is predicted by Equation 3-1; the actual data points (run-averages) are also shown. The SO_2 removal is linear with adipic acid concentration for the operating range tested, increasing from 49 percent at 1000 ppm adipic acid to 61 percent at 2000 ppm, and to 73 percent at 3000 ppm. Even if the linear relationship

Table 3-1

RESULTS OF LIME/ADIPIC ACID FACTORIAL TESTS USING THE VENTURI ONLY

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)	L/G (gal/mcf)	Press. Drop (in H_2O)	Venturi Inlet Liquor pH		Venturi Outlet Liq. pH	Ave. $AD^=$ Conc (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration (ppm)		Percent SO_2 Removal		Replicate
					Range	Ave.				Range	Ave.	Range	Ave.	
VAA101	20,000	587	37	8.8	5.99-6.14	6.06	4.60	1700	2925	2700-2800	2762	58-60	59.4	
VAA102	32,500	587	22	8.9	5.98-6.15	6.07	4.60	1545	2441	2360-2760	2585	54-59	56.2	G
VAA103	27,500	589	26	8.8	6.13-6.24	6.19	4.90	1637	2352	2700-2880	2799	57-61	59.4	
VAA104	33,000	591	22	9.0	6.03-6.23	6.13	4.50	1602	2396	2480-2680	2596	54-57	55.8	G
VAA105	20,000	587	37	9.1	5.63-6.16	5.88	4.90	2412	2086	2840-3040	2943	62-65	63.7	
VAA106	32,500	590	22	9.1	5.62-6.10	5.87	4.60	2153	2529	2480-2640	2562	54-58	55.8	
VAA107	27,500	583	26	9.0	5.46-6.15	5.87	4.60	2076	2529	2720-3000	2858	55-62	58.4	
VAA108	20,000	597	37	9.0	6.00-6.48	6.20	5.40	2821	2860	1680-1800	1742	72-78	75.5	H
VAA109	35,000	590	21	9.0	5.91-6.34	6.09	-	3079	2308	1720-2000	1863	69-75	71.2	
VAA110	27,500	597	26	8.8	5.74-6.65	6.06	5.30	3172	3191	1600-1800	1666	73-77	74.8	
VAA111	20,000	598	37	9.0	5.87-6.15	5.99	5.40	3043	3077	1680-1800	1771	73-74	73.8	H
VAA112(1)	20,000	593	37	9.0	5.47-6.50	5.97	5.24	4090→750	3132	1760-2480	2220	74→46	-	
VAA113	20,000	590	37	8.9	5.82-6.43	6.04	5.17	848	2130	2120-2280	2224	45-49	46.8	
VAA114(1)	20,000	591	37	8.9	5.35-6.60	6.02	5.80	770→140(2)	1238	1880-2480	2166	47→27	-	

(1) Adipic acid depletion tests.
(2) Open-loop operation.

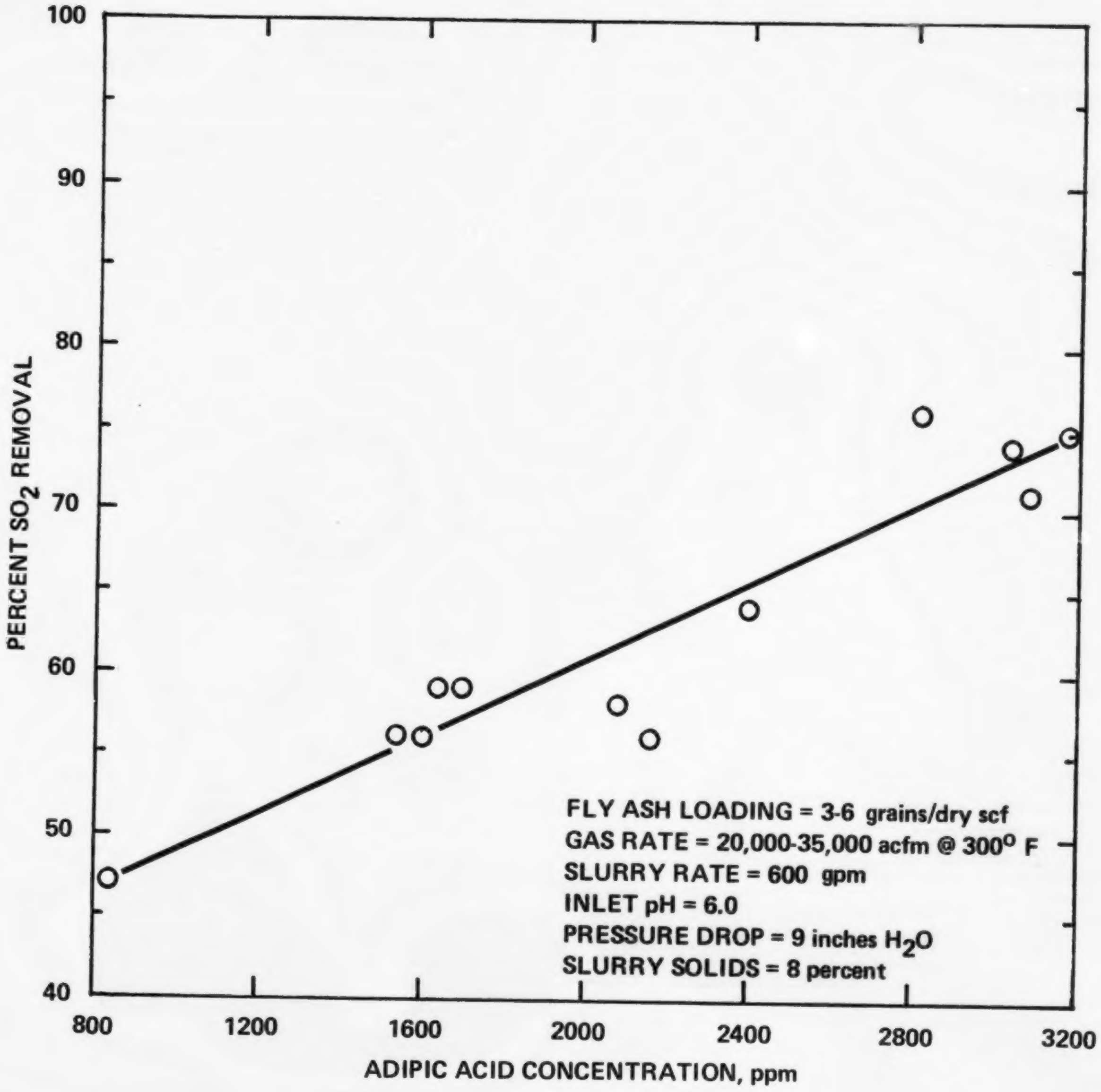


Figure 3-1. EFFECT OF ADIPIC ACID CONCENTRATION ON SO_2 REMOVAL IN THE VENTURI WITH LIME SCRUBBING

continues for higher adipic acid concentrations (an optimistic assumption), at least 4500 ppm of adipic acid would be required to achieve 90 percent SO_2 removal, approaching the gas-phase mass-transfer control limit of 94 percent SO_2 removal established by early Shawnee soda-ash tests on the venturi (Reference 5).

Figure 3-2 is a plot of venturi hourly SO_2 removal data as a function of adipic acid concentration during the adipic acid depletion tests (Runs VAA112 and VAA114). A comparison of Figures 3-1 and 3-2 shows the SO_2 removals to be in good agreement. Extrapolation of Figure 3-2 to zero adipic acid concentration gives 25 to 30 percent SO_2 removal, which is typical for the Shawnee venturi when no adipic acid is used.

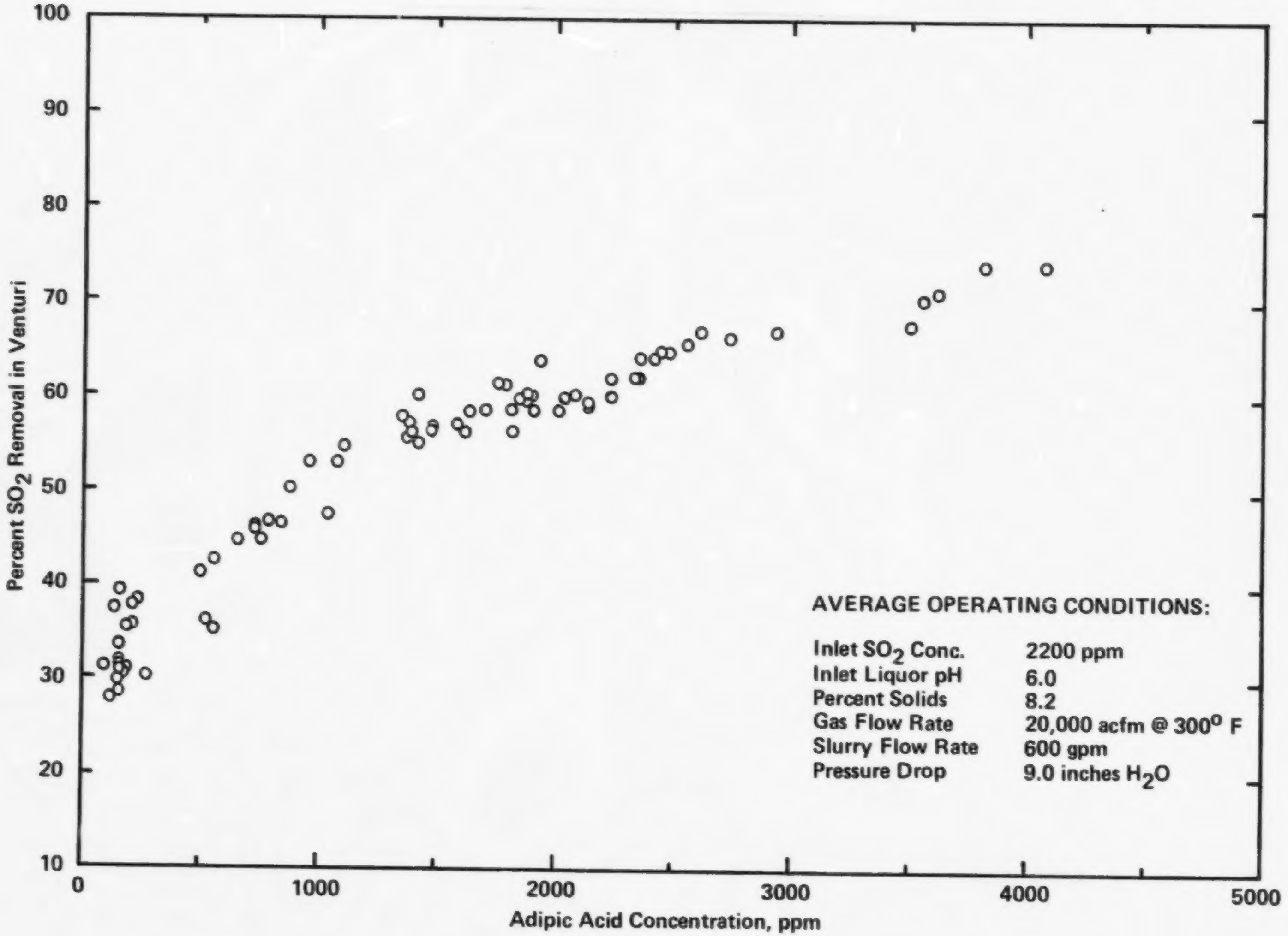


Figure 3-2. PERCENT SO_2 REMOVAL vs. ADIPIC ACID CONCENTRATION DURING ADIPIC ACID DEPLETION LIME RUNS VAA112 AND VAA114

Section 4

FACTORIAL TESTS ON THE VENTURI/SPRAY TOWER SYSTEM WITH LIMESTONE/ADIPIC ACID SLURRY

In the period from July 6 through October 3, 1979, a series of 54 factorial tests were conducted on the venturi/spray tower system with limestone/adipic acid slurry. These tests were made to investigate the effects on SO_2 removal of the following variables:

- Adipic acid concentration in liquor
- Gas flow rate and venturi/spray tower slurry flow rates (liquid-to-gas ratios)

Figure 4-1 shows the flow configuration of the two-loop venturi/spray tower system used during the factorial tests. The two-loop configuration is described in more detail in Reference 1. Tests were conducted both with and without forced oxidation. The arrangement of the oxidation tank is given in Figure 2-2.

During the majority of tests, inlet flue gas SO_2 concentrations ranged from 1800 to 2800 ppm (high sulfur coal). In 7 runs, however, low sulfur coal was burned in Boiler No. 10 and the inlet SO_2 concentrations dropped to a low range of 800 to 1800 ppm. Test results for runs with high and low sulfur coals are listed separately in Tables 4-1 and 4-2, respectively.

The 54 runs were divided into three groups of testing (see footnotes of Tables 4-1 and 4-2):

(1) Venturi and spray tower two-scrubber-loop tests without forced oxidation and with open-liquor-loop operation (no clarified liquor or filtrate was returned to the system).

(2) Spray tower only operation without forced oxidation (two-scrubber-loop with venturi plug wide open at minimum slurry flow of 125 gpm).

(3) Venturi and spray tower two-scrubber-loop tests with forced oxidation in the venturi loop.

Test group (1) was made to observe the effect of fresh adipic acid on SO_2 removal, as compared to test group (3) in which the recycled (degraded)

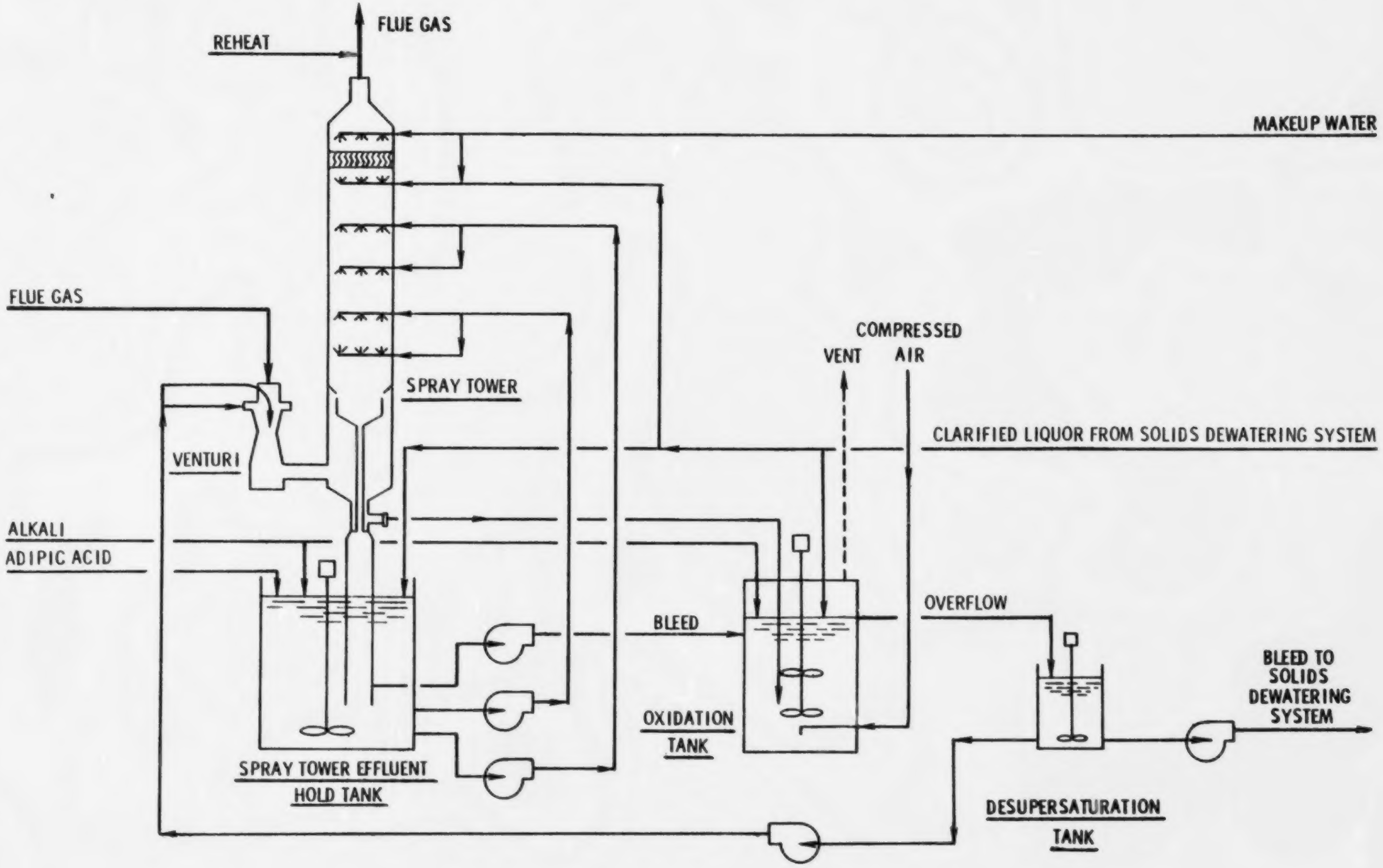

FLUE GAS
REHEAT
MAKEUP WATER
FLUE GAS
COMPRESSED AIR
VENT
SPRAY TOWER
VENTURI
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
ALKALI
ADIPIC ACID
OVERFLOW
BLEED
BLEED TO SOLIDS DEWATERING SYSTEM
OXIDATION TANK
SPRAY TOWER EFFLUENT HOLD TANK
DESUPERSATURATION TANK

Figure 4-1. FLOW DIAGRAM FOR ADIPIC ACID ENHANCED SCRUBBING IN THE VENTURI/SPRAY TOWER SYSTEM WITH TWO SCRUBBER LOOPS AND FORCED OXIDATION

Table 4-1

RESULTS OF LIMESTONE/ADIPIC ACID FACTORIAL TESTING ON THE VENTURI/SPRAY TOWER SYSTEM USING HIGH SULFUR COAL

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)		ST Gas Velocity (ft/sec)	ST L/G (gal/mcf)	Pressure Drop (inch H_2O)		Inlet Liquor pH		ST Outlet Liquor pH	Ave. ST $AD^=$ Conc. (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration (ppm)		Percent SO_2 Removal		Replicate	Notes
		Venturi	ST(1)			Venturi	Total	Venturi	ST				Range	Ave.	Range	Ave.		
VAA-001	35,000	591	1263	9.4	45	9.0	14.3	5.84	6.27	-	309	222	2320-2520	2410	91-93	92.0		(2)
VAA-002	35,000	586	1606	9.4	57	9.0	14.8	5.73	6.37	-	343	178	2280-2440	2327	94-97	95.5		(2)
VAA-003	35,000	586	1596	9.4	57	8.5	14.1	5.71	6.33	6.3	595	178	2240-2680	2482	96-98	97.5		(2)
VAA-004	35,000	586	1221	9.4	44	8.9	14.1	5.89	6.37	6.2	588	228	2400-2560	2504	95-96	95.3		(2)
VAA-005	27,500	122	1203	7.4	55	2.7	5.8	5.78	5.86	5.5	693	1464	2240-2540	2357	84-90	86.4	A	(3)
VAA-006	35,000	125	1204	9.4	43	4.2	9.1	5.55	5.97	5.8	597	888	2520-2640	2557	79-82	80.4	G	(3)
VAA-007	27,500	125	1231	7.4	56	3.1	6.1	5.81	6.01	5.8	608	977	2320-2600	2451	87-91	89.1	A	(3)
VAA-008	20,000	123	1206	5.4	75	1.5	3.2	5.79	5.90	5.7	601	1598	2000-2120	2062	94-96	94.7	H	(3)
VAA-009	20,000	122	1590	5.4	99	2.1	2.9	5.90	5.91	5.7	646	1420	2000-2120	2097	98	98.0		(3)
VAA-010	35,000	125	799	9.4	29	4.6	9.6	5.33	5.84	5.6	594	1420	1820-2320	2216	72-77	74.3		(3)
VAA-311	35,000	593	788	9.4	28	9.0	14.8	5.03	5.92	5.7	620	1627	2360-2480	2428	77-80	78.7		(4)
VAA-312	20,000	600	1550	5.4	97	8.7	10.6	5.35	5.91	5.9	693	1997	2120-2320	2186	97-98	97.6		(4)
VAA-313	27,500	595	1188	7.4	54	9.4	13.1	5.09	5.90	5.6	658	2325	2200-2280	2239	92-94	93.5	B	(4)
VAA-314	35,000	598	1190	9.4	43	9.0	15.1	5.23	5.82	5.6	653	2574	2220-2300	2263	90-91	90.6		(4)
VAA-315	20,000	594	1191	5.4	74	9.0	10.6	5.42	5.81	5.6	723	2529	2200-2280	2238	96-97	96.5		(4)
VAA-316	27,500	590	1188	7.4	54	8.9	12.4	4.63	5.92	5.6	665	2663	2120-2320	2207	91-93	92.3	B	(4)
VAA-017	35,000	591	1191	9.4	43	9.0	14.6	4.94	6.01	5.8	1012	1213	2400-2480	2429	89-92	91.5		(2)
VAA-018	35,000	596	1596	9.4	57	9.0	15.3	4.66	6.08	5.7	1037	337	2320-2480	2389	94-97	95.6		(2)
VAA-019	35,000	596	1598	9.4	57	8.9	14.5	5.34	6.17	6.1	1438	182	2000-2320	2167	98-99	98.2		(2)
VAA-020	35,000	598	1203	9.4	43	9.2	14.9	4.56	6.09	5.9	1177	275	2080-2240	2170	94-98	94.6		(2)
VAA-021	27,500	123	1192	7.4	54	2.2	6.3	4.76	5.54	5.3	1247	2535	2400-2520	2484	88-91	89.0	C	(3)
VAA-022	20,000	120	1198	5.4	74	1.5	3.0	5.08	5.61	5.4	1354	2840	2200-2300	2269	94-95	94.7		(3)
VAA-023	35,000	122	1196	9.4	43	3.5	8.8	5.06	5.77	5.5	1294	3195	1800-2160	1973	89-93	91.0		(3)
VAA-024	27,500	120	1198	7.4	54	2.3	5.2	4.93	5.56	-	1270	3816	2240-2360	2310	89-91	89.9	C	(3)
VAA-025	35,000	122	794	9.4	28	3.0	8.7	4.75	5.51	5.3	1288	2814	2400-2680	2587	73-76	73.9		(3)
VAA-026	20,000	122	1599	5.4	99	1.4	3.7	5.21	5.67	5.5	1430	2769	1880-2000	1942	96-98	96.7		(3)
VAA-027	20,000	599	1603	5.4	100	8.9	10.5	5.19	6.04	5.8	1363	612	2040-2120	2057	98-99	98.9		(4)
VAA-028	35,000	599	795	9.4	28	9.3	15.1	4.80	5.38	5.0	1459	1917	2260-2520	2426	86-90	86.9		(4)
VAA-029	27,500	598	1197	7.4	54	9.0	12.3	4.83	5.86	5.6	1211	941	2240-2480	2392	96-97	96.4	D	(4)
VAA-030	20,000	597	1205	5.4	75	9.0	10.4	4.99	5.98	5.7	1396	905	2200-2440	2277	98-99	98.4		(4)
VAA-031	35,000	594	1175	9.4	43	9.0	14.7	4.88	5.65	5.4	1477	1702	2360-2540	2951	94-95	94.4		(4)
VAA-032	27,500	598	1192	7.4	54	9.0	12.5	4.81	5.70	5.5	1209	1775	2320-2520	2411	96-97	96.8	D	(4)
VAA-035	35,000	588	795	9.4	28	8.9	14.9	4.59	5.59	5.4	2054	3343	2240-2400	2358	88-92	89.6		(2)
VAA-036	35,000	125	796	9.4	28	4.0	9.1	4.77	5.55	5.3	2238	4371	2360-2500	2455	82-84	83.3		(3)
VAA-037	20,000	124	1596	5.4	100	1.4	3.3	5.00	5.58	5.4	2139	4438	2440-2640	2547	98	98.0		(3)
VAA-038	35,000	118	1200	9.4	43	3.4	8.6	4.62	5.75	5.4	1802	3787	2200-2360	2259	88-90	88.2	E	(3)
VAA-039	27,500	125	1197	7.4	54	2.3	5.4	4.89	5.58	5.4	2081	4438	2400-2480	2453	90-92	90.9		(3)
VAA-040	20,000	125	1195	5.4	75	1.5	3.0	5.05	5.64	5.3	1830	4526	2320-2520	2438	95-96	95.9		(3)
VAA-041	35,000	126	1194	9.4	43	3.7	9.1	4.87	5.65	5.3	1773	4171	2680-2800	2697	85-86	85.1	E	(3)
VAA-042	35,000	126	400	9.4	14	3.7	8.8	4.79	5.62	5.4	1973	3861	2400-2560	2456	59-61	60.5		(3)
VAA-043	35,000	589	797	9.4	28	9.0	15.1	4.71	5.24	5.1	2059	3994	2200-2400	2317	89-92	90.2	F	(4)
VAA-044	35,000	587	1209	9.4	43	9.0	15.1	4.96	5.65	5.5	1853	3595	2080-2200	2118	95-96	95.8		(4)
VAA-045	27,500	589	1205	7.4	55	8.9	13.1	5.07	5.75	5.3	2124	4083	2200-2520	2321	97-98	97.4		(4)
VAA-046	20,000	590	1207	5.4	75	8.8	10.7	4.89	5.33	5.1	1959	4319	2240-2280	2265	97-98	97.4		(4)
VAA-048	35,000	586	794	9.4	28	9.0	15.0	4.74	5.21	5.0	2124	5029	2320-2440	2388	86-88	87.0	F	(4)
VAA-306	35,000	124	1190	9.4	43	5.0	11.3	4.94	5.75	5.3	597	2201	2120-2440	2316	77-81	80.1	G	(3)
VAA-308	20,000	125	1190	5.4	75	1.8	3.6	5.16	5.86	5.7	591	1864	1800-2240	2011	90-93	91.7	H	(3)

Notes:

(1) Spray tower slurry flow rates are in 400 gpm increments (400, 800, 1200, and 1600 gpm). For 1600 gpm full flow rate, all 4 headers are used at 400 gpm/header. Headers are turned off starting from the bottom header for runs using less than 1600 gpm.
(2) Venturi and spray tower two-scrubber-loop tests without forced oxidation and with open-liquor-loop operation.
(3) Spray tower only operation (venturi plug wide open with 125 gpm minimum slurry flow) without forced oxidation.
(4) Venturi and spray tower two-scrubber-loop tests with forced oxidation.

Table 4-2

RESULTS OF LIMESTONE/ADIPIC ACID FACTORIAL TESTING ON THE VENTURI/SPRAY TOWER SYSTEM USING LOW SULFUR COAL

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)		ST Gas Velocity (ft/sec)	ST L/G (gal/mcf)	Pressure Drop (inch H_2O)		Inlet Liquor pH		ST Outlet Liquor pH	Ave. ST $AD^=$ Conc. (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration, (ppm)		Percent SO_2 Removal		Replicate	Notes
		Venturi	ST(1)			Venturi	Total	Venturi	ST				Range	Ave.	Range	Ave.		
VAA-305	27,500	116	1190	7.4	55	3.6	8.7	4.99	5.54	5.2	629	2130	1480-1600	1520	73-86	82.0		(2)
VAA-011	35,000	589	784	9.4	28	9.0	15.4	5.09	6.15	6.0	637	1154	800-1680	1347	89-98	94.2		(3)
VAA-012	20,000	585	1595	5.4	100	9.0	10.9	5.08	5.84	5.7	687	1625	1080-1400	1304	96-98	96.9		(3)
VAA-013	27,500	592	1191	7.4	55	9.0	12.9	5.14	6.28	5.9	665	1310	1320-1600	1458	95-97	95.9	B	(3)
VAA-014	35,000	584	1193	9.4	43	9.1	15.4	4.80	6.02	5.8	643	1287	1600-1780	1698	92-94	93.2		(3)
VAA-015	20,000	587	1198	5.4	75	9.0	11.4	5.29	5.91	5.6	704	1674	1280-1600	1420	97-98	97.4		(3)
VAA-016	27,500	586	1196	7.4	55	8.8	12.7	5.15	6.18	5.9	562	1154	1360-1680	1502	96-98	96.8	B	(3)

Notes:

(1) Spray tower slurry flow rates are in 400 gpm increments (400, 800, 1200, and 1600 gpm). For 1600 gpm full flow rate, all 4 headers are used at 400 gpm/header. Headers are turned off starting from the bottom header for runs using less than 1600 gpm.

(2) Spray tower only operation (venturi plug wide open with 125 gpm minimum slurry flow) without forced oxidation.

(3) Venturi and spray tower two-scrubber-loop tests with forced oxidation.

adipic acid from the dewatering system made up most of the adipic acid in the scrubber loops. Test group (2) was made to investigate the SO_2 removal efficiency with the spray tower alone.

The operating conditions common to all the factorial tests were:

Fly ash loading: High (3-6 grains/dry scf)
Spray tower hold tank level (20 ft diameter): 10 ft
Oxidation tank level (8 ft diameter): 18 ft
Desupersaturation tank level (10 ft diameter): 4.8 ft
Venturi slurry flow rate: 600 gpm for test groups (1) & (3)
125 gpm for test group (2)
Venturi inlet pH: Greater than 4.5 for test groups (1) & (3)
Spray tower limestone stoichiometry: 1.7 for test groups (1) & (3)
1.2 for test group (2)
Air flow rate to oxidizer: 220 scfm for test group (3) only
Venturi slurry solids conc.: 15 percent for test groups (1) & (3)
uncontrolled for test group (2)
Spray tower slurry solids conc.: 6 percent for test groups (1) & (3)
15 percent for test group (2)
Venturi pressure drop: 9 in. H_2O for test groups (1) & (3)
wide open plug for test group (2)

For a given controlled level of adipic acid concentration in the spray tower liquor, the corresponding adipic acid concentration in the venturi liquor was normally about 50 percent higher. Flue gas humidification (evaporation of water) in the venturi resulted in higher dissolved species concentrations in the venturi liquor.

RESULTS OF TESTING WITH SPRAY TOWER ONLY - TEST GROUP (2)

In the present equipment set-up at Shawnee, the flue gas cannot be by-passed around the venturi scrubber for testing with the spray tower alone. In this series of tests, a compromise was made by operating the venturi with wide-open plug and with the minimum controllable slurry flow rate of 125 gpm for flue gas cooling to protect the neoprene lining of the spray tower. This mode of operation minimizes the effect of the venturi on SO_2 removal and allows the evaluation of the spray tower performance.

Unfortunately, with adipic acid-enhanced scrubbing, a significant amount of SO_2 was removed in the venturi even though only 125 gpm of slurry was used. For example, in Run VAA022 (Table 4-1) SO_2 removal by venturi alone was 33 percent (determined by turning off the spray tower slurry flow for 30 minutes at the end of the test) at 20,000 acfm gas rate and about 2000 ppm adipic acid in venturi liquor. This removal was certainly not insignificant compared to the overall removal of 94.7 percent.

Figure 4-2 is a plot of the SO_2 removal as a function of spray tower adipic acid concentration and spray tower liquid-to-gas ratio (L/G) for runs with high inlet SO_2 concentration. The SO_2 removals in Figure 4-2 include the effect of the venturi scrubber as mentioned above. As can be seen, the SO_2 removal is sensitive to L/G and also to adipic acid concentration at low L/G.

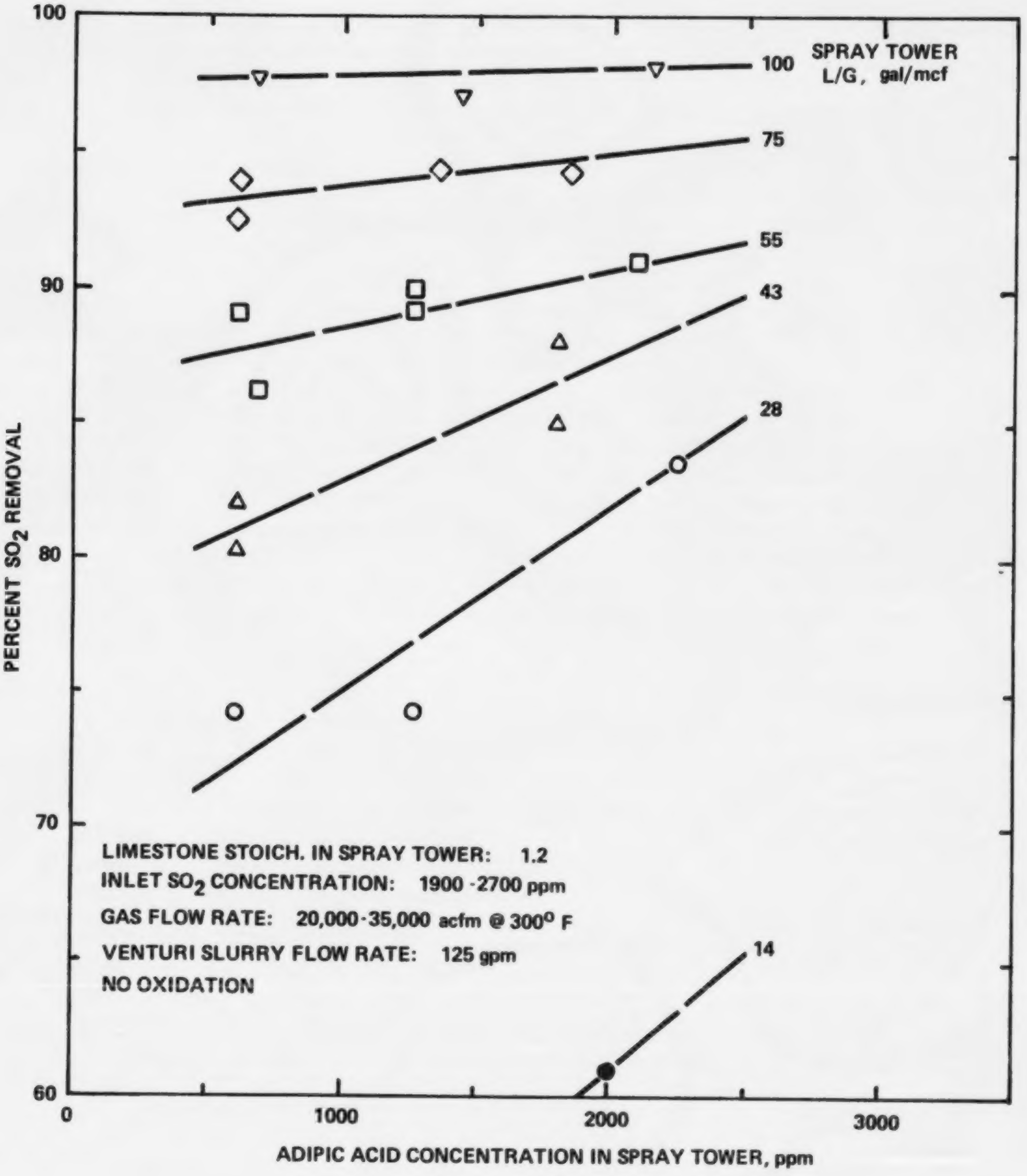


FIGURE 4-2

EFFECT OF ADIPIC ACID CONCENTRATION AND SPRAY TOWER L/G ON SO_2 REMOVAL IN THE SPRAY TOWER

Because of the significant effect on the overall SO_2 removal by the venturi, no attempt was made to correlate the data.

RESULTS OF TESTING WITH VENTURI AND SPRAY TOWER IN TWO-LOOP CONFIGURATION -- TEST GROUPS (1) & (3)

For all of these tests, the slurry flow rate to the venturi was held constant at 600 gpm and the venturi plug position was varied according to the gas rate to maintain a constant gas pressure drop of 9 inches H_2O. Variations in the spray tower slurry flow rate and gas rate resulted in a spray tower L/G range of 28 to 100 gal/mcf. The limestone stoichiometry in the spray tower was controlled at 1.7 mole Ca/mole SO_2 absorbed.

Test results indicated that there was insignificant difference in the overall SO_2 removals between runs in test group (1) made with open-liquor-loop operation without forced oxidation and runs in test group (3) made with closed-liquor-loop and forced oxidation. Open-liquor-loop tests require a much higher adipic acid addition rate, and since no spent liquor is returned to the scrubbing loops from the dewatering system the adipic acid in the scrubbing loops should be relatively "fresh" and not degraded to any significant degree. Closed-liquor-loop tests combined with forced oxidation, on the other hand, provide the highest concentration of "degraded" adipic acid. In theory, however, the lower-carbon carboxylic acids (such as valeric acid, butyric acid, etc.) produced by the degradation and decomposition of adipic acid should also provide effective buffer capacity (Reference 3).

Figure 4-3 shows the SO_2 removal for runs with high inlet SO_2 concentrations as a function of adipic acid concentration and L/G in the spray tower. For the spray tower L/G greater than 43 gal/mcf, SO_2 removals were all above 90 percent for spray tower adipic acid concentrations as low as 300 ppm. As in the spray tower-only operation (Figure 4-2), SO_2 removal was sensitive to both L/G and adipic acid concentration in the spray tower at a low L/G of 28 gal/mcf.

No correlation of the data was attempted because of the high SO_2 removals (greater than 90 percent) for a wide range of L/G from 43 to 100 gal/mcf (Figure 4-3).

For runs with low inlet SO_2 concentrations (Table 4-2), SO_2 removals were higher at 93 to 97 percent at 600 ppm adipic acid concentration and L/G of 43 to 100 gal/mcf in the spray tower.

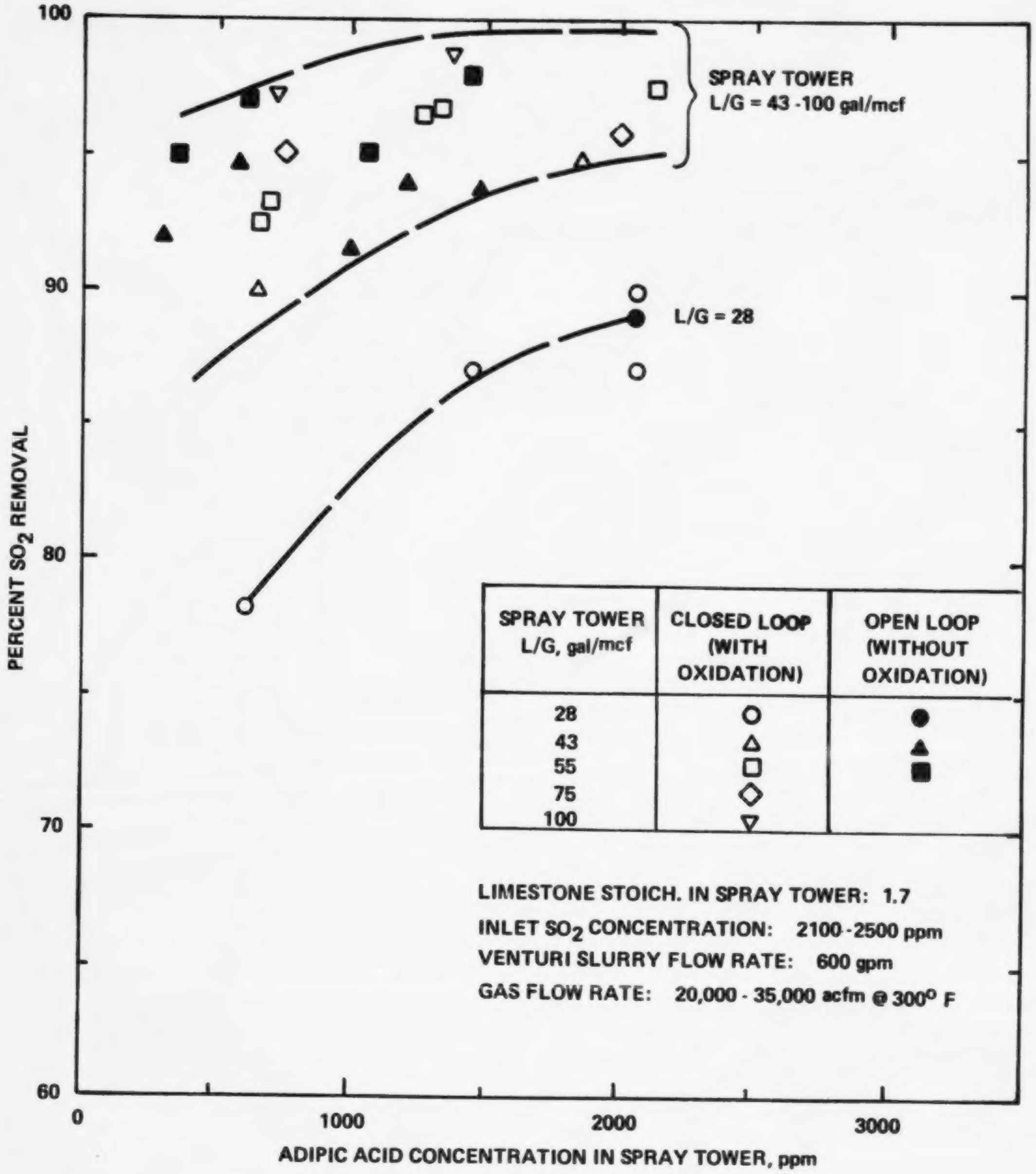


FIGURE 4-3

EFFECT OF ADIPIC ACID CONCENTRATION AND L/G IN SPRAY TOWER ON SO_2 REMOVAL IN THE VENTURI/SPRAY TOWER

Section 5

SIMULATION OF WIDOWS CREEK FULL-SCALE FORCED OXIDATION ON THE VENTURI/SPRAY TOWER SYSTEM

Beginning in mid 1979, TVA conducted a full-scale forced-oxidation demonstration in the scrubber Module D of boiler No. 8 at their Widows Creek Steam Plant in Bridgeport, Alabama. The scrubber system consists of a venturi followed by a grid tower and operates in a two-scrubber-loop mode. Some backmixing of slurry from the venturi loop (oxidation loop) to the grid tower loop is expected by the slurry entrainment in the flue gas.

In order to obtain backup data for the full-scale demonstration, a six-week test was conducted at Shawnee from mid-February through March 1979 using the venturi/spray tower system with an artificial backmix stream to simulate the full-scale operation. A flow diagram of the set-up at Shawnee is given in Figure 5-1. The details of the oxidation tank are shown in Figure 2-2.

One of the main purposes of the Widows Creek simulation was to test the effect of Widows Creek limestone (Stoneman) and Widows Creek coal (Pittsburgh Midway) on sulfite oxidation, as compared to the oxidation normally observed at Shawnee when Shawnee limestone (Fredonia) and Shawnee coal were used. A wide range of mid- to high-sulfur coals are burned at Shawnee, including Pittsburgh Midway coal.

Both the laboratory limestone reactivity test results (Reference 9) and the results of this test series indicated that there was a slight difference between Stoneman and Fredonia limestone. All tests showed Stoneman limestone to be somewhat less reactive than Fredonia for SO_2 removal. The particle size distribution of 87 percent less than 325 mesh for Stoneman limestone (ground at Shawnee) was slightly more coarse than the typical Fredonia limestone of 91 percent less than 325 mesh used at the test facility.

No difference in sulfite oxidation, with respect to catalytic effect from trace elements, could be detected from the use of Widows Creek coal or limestone. The inlet SO_2 concentration were very consistent and fell between 2300 and 3100 ppm when Widows Creek coal was used.

DISCUSSION OF TEST RESULTS

Table 5-1 summarizes the major test conditions and the average results for the 5 runs made in this test series. All runs were made without adipic acid additive.

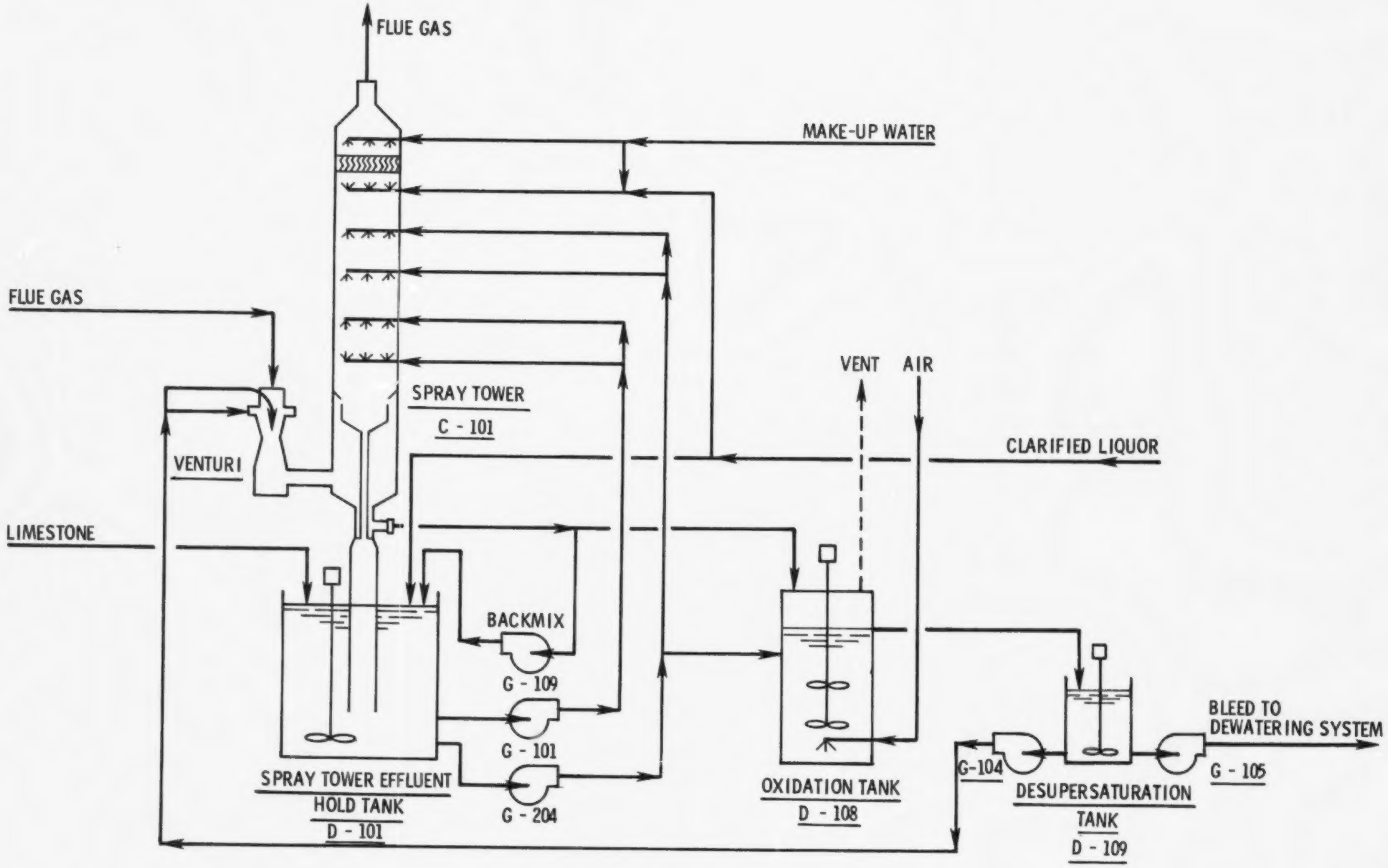

FLUE GAS
MAKE-UP WATER
FLUE GAS
VENTURI
SPRAY TOWER
C - 101
VENT
AIR
CLARIFIED LIQUOR
LIMESTONE
BACKMIX
G - 109
G - 101
G - 204
SPRAY TOWER EFFLUENT
HOLD TANK
D - 101
OXIDATION TANK
D - 108
G-104
DESUPERSATURATION
TANK
D - 109
BLEED TO
DEWATERING SYSTEM
G - 105

Figure 5-1. FLOW CONFIGURATION FOR WIDOWS CREEK SIMULATION ON THE VENTURI/SPRAY TOWER WITH BACKMIX

Table 5-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE TESTS WITH TWO-SCRUBBER-LOOP FORCED OXIDATION AND BACKMIX -WIDOWS CREEK SIMULATION-

Major Test Conditions(1)	827-1A	828-1A	828-1B	829-1A	830-1A
Fly ash loading	High	High	High	High	High
Gas rate, acfm @ 300°F	35,000	35,000	35,000	35,000	35,000
Venturi liquor rate, gpm	600	600	600	600	600
Spray tower liquor rate, gpm	1500	1500	1500	1500	1500
Venturi percent solids recirculated (controlled)	12	12	12	12	12
Spray tower percent solids recirculated (controlled)(2)	10	10	10	10	10
Residence time (min)/Tank level (ft):					
Spray tower EHT	9.4/6	9.4/6	9.4/6	9.4/6	9.4/6
Oxidation tank	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18
Desupersaturation tank	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8
Venturi inlet pH (controlled)	5.5	5.5	(3)	5.5	6.0
Air rate to oxidizer, scfm	250	180	180	145	250
Backmix rate, gpm	65	65	65	65	29
Run-Average Results					
Start-of-Run date	2/14/79	3/12/79	3/13/79	3/16/79	3/22/79
Onstream hours	345	29	48	106	181
Percent SO_2 removal	75	83	59	79	89
Inlet SO2 concentration, ppm	2800	2600	2750	2850	2600
Spray tower percent solids recirculated	10.7	10.6	10.6	10.5	9.9
Venturi inlet pH	5.5	5.55	5.25	5.55	6.0
Spray Tower inlet pH	5.6	5.65	5.0	5.75	6.1
Spray tower limestone stoichiometric ratio	2.0	2.6	2.1	1.7	2.4
Spray tower inlet liquor gypsum saturation, %	100	95	115	90	90
Spray tower sulfite oxidation	72	70	93	65	55
Overall sulfite oxidation, %	98	98	98.5	87	94
Overall limestone utilization, %	60	46	68	70	56
Venturi inlet liquor gypsum saturation, %	100	95	95	95	100
Venturi inlet liquor $SO_3^{=}$ concentration, ppm	65	115	45	170	85
Air stoichiometry, lb-atoms O/lb mole SO_2 absorbed	2.1	1.45	1.95	1.15	1.90
Filter cake solids, wt%	85	84	84	79	79
Mist eliminator restriction, %	-	-	0	-	0

(1) All runs used Widows Creek limestone (Stoneman) and coal (Pittsburgh Midway). Pressure drop across the venturi was 9 in. H_2O for all runs.

(2) Controlled by splitting the returned clarified liquor to the oxidation tank and spray tower EHT.

(3) Venturi stoichiometry controlled at 1.3.

Run 827-1A. One of the main concerns was that the differences in trace element contents in limestone and coal used at Widows Creek might cause poor or no oxidation of sulfite which has been rather easily achieved at Shawnee. To simulate operating conditions a portion (65 gpm) of the venturi outlet slurry was returned to the spray tower slurry recirculation system (see Figure 5-1).

A high air stoichiometry was used to assure good oxidation by providing 250 scfm of air to the oxidizer. The resulting air stoichiometry was 2.1 atoms oxygen/mole SO_2 absorbed. Overall oxidation under these conditions was 98 percent with 85 percent solids in the filter cake.

Table 5-1 shows that the SO_2 removal averaged only 75 percent at an inlet SO_2 level of 2800 ppm. This is lower than what would be expected under conditions of no backmixing, but is not unexpected considering the lower pH of 5.6 in the spray tower inlet slurry caused by backmixing. The closeness of the values in the venturi and spray tower loops indicate that the system approaches single loop operation with the backmix.

Run 828-1A. Since good oxidation was achieved in Run 827-1A at a rather high air stoichiometry of 2.1 atoms oxygen/mole SO_2 absorbed, air flow rate was reduced to 180 scfm in this run. Sulfite oxidation remained high at 98 percent at an air stoichiometry of 1.45 atoms oxygen/mole SO_2 absorbed.

While oxidation remained high the limestone utilization, which had been low in Run 827-1A (60 percent), became excessively low (46 percent) during Run 828-1A. To achieve better utilization and reduce limestone consumption, a new run was made (Run 828-1B) which operated under limestone stoichiometry control of 1.3 instead of pH control of 5.5 in the venturi inlet slurry.

During Run 828-1A the SO_2 removal increased to 83 percent largely because of a lower SO_2 inlet concentration of 2600 ppm and higher venturi and spray tower limestone stoichiometries.

Run 828-1B. This run was a continuation of Run 828-1A with the limestone stoichiometry in the venturi inlet controlled (at 1.3) rather than the pH. The result was higher limestone utilization (68 percent) but lower pH's (5.3 and 5.0 in the venturi and spray tower inlet, respectively) and a subsequently lower SO_2 removal of 59 percent. The lower pH combined with the lower SO_2 removal and the correspondingly higher air stoichiometry of 1.95 assured a good oxidation of 98.5 percent.

Run 829-1A. In this run the control was returned to the venturi inlet pH at 5.5 and the air rate to the oxidizer was further reduced to 145 scfm. The resulting air stoichiometry dropped to 1.15 which in turn reduced overall sulfite oxidation to 87 percent. This confirms earlier tests under standard Shawnee operating conditions which also showed the break point for oxidation (greater than 90 percent) to lie somewhere between 1.1 and 1.5 air stoichiometry.

Despite raising the venturi inlet pH back to 5.5 the overall limestone utilization remained about the same as Run 828-1B at 70 percent. SO_2 removal averaged 79 percent at an average inlet SO_2 concentration of 2850 ppm.

Run 830-1A. In this run the venturi inlet pH was raised to 6.0 to improve the SO_2 removal, the air rate was raised back to 250 scfm (as in Run 827-1A) to insure good oxidation at the higher pH, and the backmix rate was reduced to 29 gpm.

The combination of higher pH and reduced backmix had the predictable effect of raising SO_2 removal to 89 percent at an inlet SO_2 concentration of 2600 ppm. Because of the higher pH (cf. Run 827-1A) sulfite oxidation was only 94 percent. Overall limestone utilization was slightly lower at 56 percent.

Limestone Blinding and $CaSO_3$ Supersaturation. Upon examination of the liquor analytical data, it was found that sulfite concentrations were excessively high for Run 828-1B, averaging 2935 ppm and 2500 ppm at the spray tower inlet and outlet, respectively. For other runs in the test series, sulfite concentrations ranged only from 250 to 700 ppm both at the spray tower inlet and outlet. The apparent $CaSO_3$ supersaturation in the spray tower during Run 828-1B was probably caused by the backmix stream which depleted the $CaSO_3$ solids in the spray tower loop. This was evidenced by the high spray tower solids sulfite oxidation of 93 percent for Run 828-1B, the highest in the test series. $CaSO_3$ must nevertheless crystallize in the spray tower, causing limestone blinding by $CaSO_3$ and the subsequent pH drop in the spray tower (spray tower inlet pH lower than venturi inlet pH in this case) and poor limestone utilization.

SUMMARY

The following is a summary of the test results:

- No apparent difference in sulfite oxidation could be detected from the use of Widows Creek Stoneman limestone or Pittsburgh Midway coal. Sulfite oxidation of 98 percent was achieved at 5.5 pH and an air stoichiometry as low as 1.45 atoms oxygen/mole SO_2 absorbed.
- Backmixing of slurry from the venturi loop (oxidation loop) to the spray tower loop depletes $CaSO_3$ solids in the spray tower and could cause $CaSO_3$ supersaturation and limestone blinding in the spray tower.
- Backmixing, in effect, creates an operating mode approaching that of a single-loop operation.

Section 6

LIME/ADIPIC ACID TESTING ON THE TCA WITH ONE TANK AND WITHOUT FORCED OXIDATION

In the last report (Reference 1) results of a single adipic acid-enhanced lime test (made in July 1978) without forced oxidation were presented. The concentration of adipic acid used in that run (Run 977-2A) was 420 ppm. In that run the SO_2 removal improved from 67 percent for the base case run without adipic acid (Run 976-2B) to 80 percent, with a decrease in SO_2 concentration from 2950 to 2700 ppm. Both runs were made at a scrubber inlet pH of 7.0.

Additional adipic acid-enhanced lime tests with one tank and without forced oxidation were conducted in February and March 1979. Adipic acid concentrations up to 1300 ppm were used in these tests.

DISCUSSION OF TEST RESULTS

Table 6-1 presents a summary of test results for three lime tests with adipic acid addition along with a base case lime run without adipic acid. All runs were made with a single tank and without forced oxidation. Controlled run conditions common to all of the tests were:

Fly ash loading: High (3 to 6 grains/dry scf)
Gas flow rate: 30,000 acfm @ 300°F
Slurry solids concentration: 8 percent
Scrubber inlet pH: 7.0
Effluent hold tank level (7 ft. diameter): 17 ft.
Total static bed height: 15 inches of 1-7/8 inch diameter nitrile foam spheres
Makeup lime addition point: Effluent hold tank

Effect of Adipic Acid Concentration. For Runs 978-2A, 979-2A, and 980-2A, average adipic acid concentrations were 0, 615, and 1305 ppm, respectively, with the corresponding SO_2 removals of 83, 93, and 97.5 percent. Inlet SO_2 concentrations were 2350, 2900, and 2750 ppm, respectively. Thus, with 600 to 1300 ppm of adipic acid SO_2 removal improved by 10 to 15 percentage points over the base case.

6-1

Table 6-1

RESULTS OF TCA LIME/ADIPIC ACID TESTS WITHOUT FORCED OXIDATION

Major Test Conditions(1)	978-2A	979-2A	980-2A	981-2A
Fly ash loading	High	High	High	High
Adipic acid concentration, ppm	0	600	1200	1200
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000
Liquor rate, gpm	1200	1200	1200	900
Percent solids recirculated	8	8	8	8
TCA inlet pH controlled at	7.0	7.0	7.0	7.0
EHT (D-204) residence time, min	4.1	4.1	4.1	5.4
EHT (D-204) level, ft	17	17	17	17
Lime addition point	EHT	EHT	EHT	EHT
Run-Average Results				
Start-of-Run date	2/8/79	2/13/79	2/22/79	3/12/79
Onstream hours	116	177	247	70
Percent SO_2 removal	83	93	97.5	92
Inlet SO_2 concentration, ppm	2350	2900	2750	2600
SO_2 make-per-pass, m-moles/l	10.3	14.3	14.3	17.0
TCA inlet liquor percent gypsum saturation	125	90	75	100
Percent sulfite oxidation	21	14	10	9
Percent lime utilization	92	92	88	90
TCA inlet pH	7.0	7.2	6.95	6.95
Adipic acid concentration, ppm	0	615	1305	1235
Mist eliminator restriction, percent	-	-	-	0

Note: (1) All runs were made with 3 beds and 5 inches per bed of 1-7/8 inch diameter nitrile foam spheres. Internal D-204 downcomer was used in all runs.

Effect of Slurry Flow Rate. Run 981-2A was made with a reduced slurry flow rate of 900 gpm, compared to 1200 gpm for Run 980-2A. Other operating conditions were the same for the two runs. As a result of the reduced slurry flow rate, the liquid-to-gas ratio decreased from 50 to 37 gal/mcf and the effluent hold tank residence time increased from 4.1 to 5.4 minutes. As expected, SO_2 removal dropped from 97.5 percent at 2750 ppm inlet SO_2 concentration to 92 percent at 2600 ppm inlet SO_2 concentration. SO_2 make-per-pass* increased from 14.3 to 17.0 m-moles/liter. Average adipic acid concentration was slightly lower at 1235 ppm for Run 981-2A.

SUMMARY

At a scrubber inlet pH of 7 and a liquid-to-gas ratio of 50 gal/mcf, adipic acid-enhanced lime scrubbing improved the SO_2 removal by 10 to 15 percentage points with 600 to 1300 ppm adipic acid over the base case SO_2 removal of 83 percent without adipic acid. Reducing the liquid-to-gas ratio to 37 gal/mcf at 1235 ppm adipic acid and 7.0 scrubber inlet pH still yielded a good SO_2 removal of 92 percent.

---

*Defined as milligram-moles of SO_2 absorbed from the flue gas per liter of liquor as the recirculated slurry makes one pass through the scrubber.

Section 7

LIME/ADIPIC ACID TESTING ON THE TCA WITH ONE TANK AND FORCED OXIDATION

Within-scrubber-loop forced oxidation in a single-loop scrubbing system would not be expected to give good SO_2 removal for a lime scrubber because of the oxidation of the major scrubbing species, sulfite ion, into non-reactive sulfate ion. With adipic acid addition, however, satisfactory SO_2 removal should be possible because of the calcium adipate buffer. In addition, the lower pH at which a lime/adipic acid system operates should facilitate sulfite oxidation. All runs in this series utilized 1200 ppm adipic acid.

SYSTEM DESCRIPTION

The flow diagram for the adipic acid-enhanced scrubbing in the single-loop TCA system with one tank and forced oxidation is shown in Figure 7-1. Adipic acid was added as a dry powder to the effluent hold tank (oxidation tank). Air was sparged into the tank through a simple 3-inch pipe ell. The arrangement of the oxidation tank is shown in Figure 7-2 which is drawn roughly to scale. The oxidation tank is 7 ft. in diameter and was operated at 17 ft. level in all runs in this test series. The agitator in the oxidation tank has two downflow turbines. Each turbine is 55 inches in diameter and contains 4 blades. Each blade on the bottom turbine has a stabilizing fin. The bottom turbine is about 20 inches above the air sparger. The agitator has a 20 Hp motor and the speed can be varied between 18 and 54 rpm. The internal scrubber downcomer was used in all runs. The clarifier and centrifuge in series were used for solids dewatering.

DISCUSSION OF TEST RESULTS

A total of 5 runs were made from mid-March through early April 1979. Table 7-1 shows a summary of major test conditions and average results for these runs. Operating conditions for all runs were the same except that the TCA inlet pH was varied from run to run in order to seek an optimum pH range for good SO_2 removal as well as good sulfite oxidation.

Run 951-2A. The first test was made at a TCA inlet pH of 7.0. The test conditions for this run were the same as Run 980-2A (Table 6-1) except that 185 scfm of air was introduced to the oxidizer during Run 951-2A. Sulfite oxidation

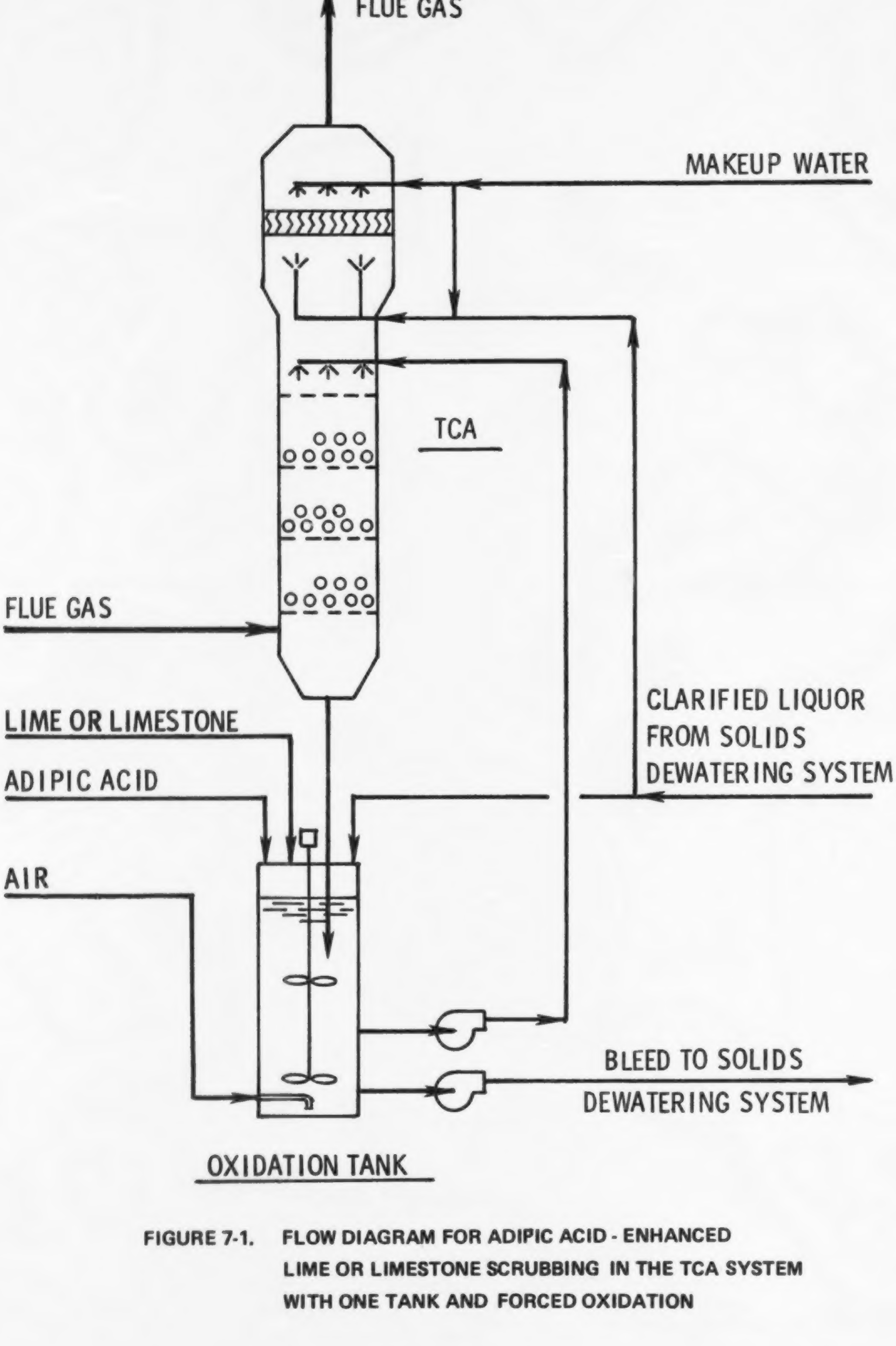

FLUE GAS
MAKEUP WATER
TCA
FLUE GAS
LIME OR LIMESTONE
ADIPIC ACID
CLARIFIED LIQUOR
FROM SOLIDS
DEWATERING SYSTEM
AIR
BLEED TO SOLIDS
DEWATERING SYSTEM
OXIDATION TANK

FIGURE 7-1. FLOW DIAGRAM FOR ADIPIC ACID - ENHANCED LIME OR LIMESTONE SCRUBBING IN THE TCA SYSTEM WITH ONE TANK AND FORCED OXIDATION

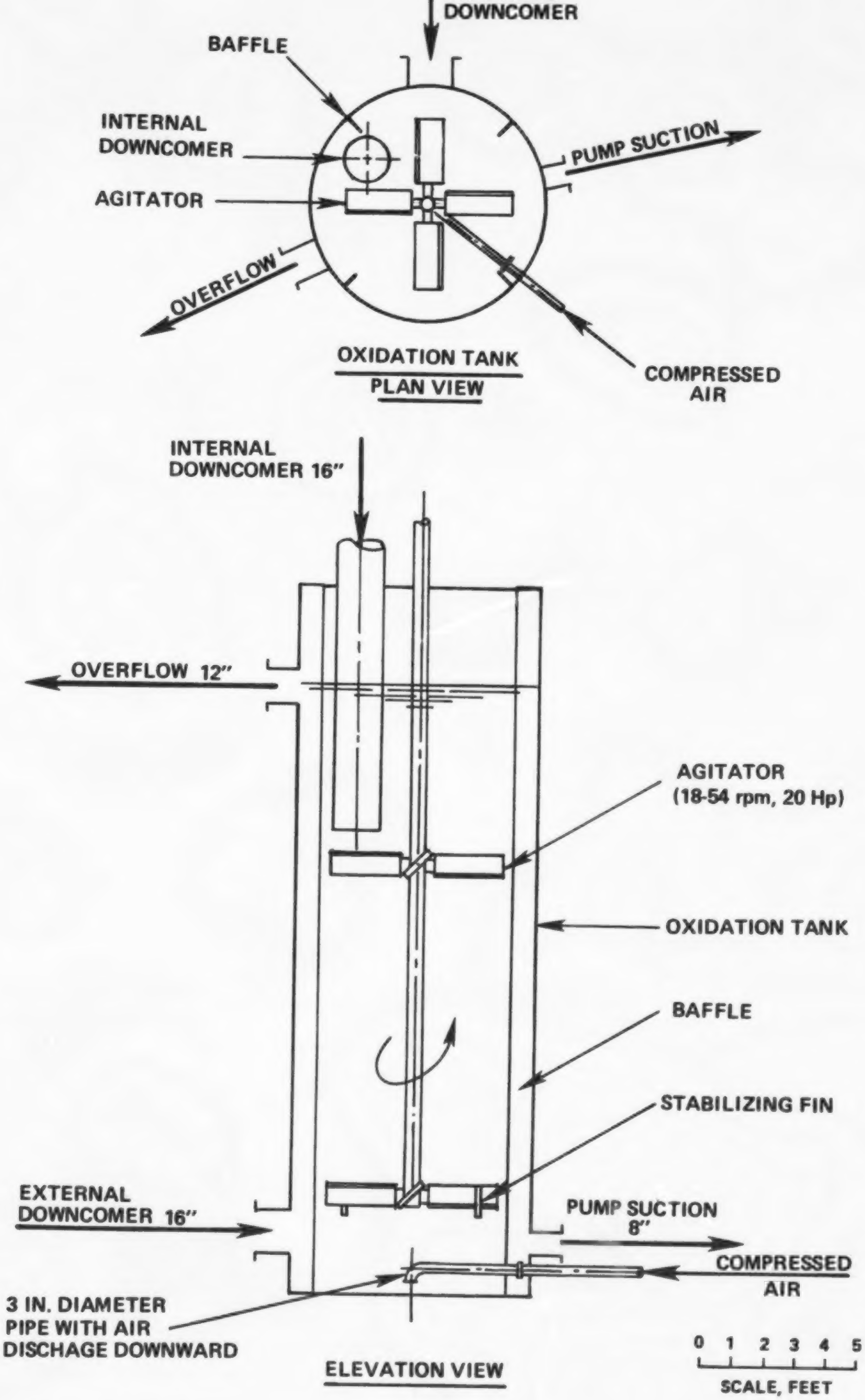

EXTERNAL DOWNCOMER
BAFFLE
INTERNAL DOWNCOMER
AGITATOR
PUMP SUCTION
OVERFLOW
COMPRESSED AIR
OXIDATION TANK
PLAN VIEW
INTERNAL DOWNCOMER 16"
OVERFLOW 12"
AGITATOR (18-54 rpm, 20 Hp)
OXIDATION TANK
BAFFLE
STABILIZING FIN
EXTERNAL DOWNCOMER 16"
PUMP SUCTION 8"
COMPRESSED AIR
3 IN. DIAMETER PIPE WITH AIR DISCHAGE DOWNWARD
ELEVATION VIEW
0 1 2 3 4 5
SCALE, FEET

Figure 7-2. ARRANGEMENT OF THE TCA OXIDATION TANK (D-204)

Table 7-1

RESULTS OF TCA LIME/ADIPIC ACID TESTS WITH FORCED OXIDATION

Major Test Conditions(1)	951-2A	951-2B	951-2C	951-2D	951-2E
Fly ash loading	High	High	High	High	High
Adipic acid concentration, ppm	1200	1200	1200	1200	1200
Gas rate, acfm @ 300^{o}F	30,000	30,000	30,000	30,000	30,000
Liquor rate, gpm	1200	1200	1200	1200	1200
Percent solids recirculated	8	8	8	8	8
TCA inlet pH (controlled)	7.0	5.5	5.0	4.5	5.0
Oxidation tank residence time (min)/Tank level (ft)	4.1/17	4.1/17	4.1/17	4.1/17	4.1/17
Lime addition point	Oxid. Tk	Oxid. Tk	Oxid. Tk	Oxid. Tk	Oxid. Tk
Air rate to oxidizer, scfm(2)	185	185	185	185	185
Run-Average Results					
Start-of-Run date	3/15/79	3/20/79	3/24/79	3/28/79	3/31/79
Onstream hours	57	102	89	64	103
Percent SO_2 removal	97	95	93	85	82
Inlet SO_2 concentration, ppm	2800	2550	2550	2150	2400
SO_2 make-per-pass, m-moles/l	14.4	12.8	12.5	9.7	10.4
TCA inlet liquor $SO_3^=$ concentration, ppm	165	245	230	80	100
TCA inlet liquor percent gypsum, saturation	105	115	110	110	105
Slurry percent solids	8.8	8.0	8.2	7.8	7.8
Percent sulfite oxidation	47	88	95	96	98
Percent lime utilization	93	96	97	98	97
TCA inlet pH	7.0	5.5	5.0	4.6	4.95
Adipic acid concentration	1215	1230	1270	1260	1330
Centrifuge cake solids, wt%	61	74	74	71	72
Air stoichiometry, atoms O/mole SO_2 absorbed	1.40	1.55	1.60	2.05	1.95
Mist eliminator restriction, %	-	-	-	0	-

Notes: (1) All runs were made with 3 beds and 5 inches/bed of 1-7/8 inch diameter nitrile foam spheres. Internal D-204 downcomer was used in all runs.

(2) A single 3-inch diameter pipe was used. Air discharged downward at center of oxidation tank 5 inches from tank bottom.

increased from 10 percent for Run 980-2A without air to only 47 percent for Run 951-2A with an air stoichiometry of 1.4 atoms oxygen/mole SO_2 absorbed. This resulted in a slight increase in centrifuge cake solids of 57 percent to 61 percent. The SO_2 removal remained high at 97 percent at 2800 ppm inlet SO_2 concentration.

Run 951-2B. To improve sulfite oxidation, the TCA inlet pH was lowered to 5.5. Oxidation increased to 88 percent while the SO_2 removal remained high at 95 percent with 2550 ppm inlet SO_2 concentration. The centrifuge cake solids content increased further to 74 percent. The high SO_2 removal obtained in this run was later attributed partly to high scrubber pressure drop caused by calcium sulfite scaling as will be explained later.

Run 951-2C. The scrubber inlet pH was further reduced to 5.0 in this run. Oxidation continued to increase to 95 percent, while SO_2 removal again remained high at 93 percent at 2550 ppm inlet SO_2 concentration. As in Run 951-2B, the unexpectedly high SO_2 removal was attributed partly to high scrubber pressure drop. At an inlet pH of 5.0 it is expected that calcium adipate is the main SO_2 scrubbing media with very little scrubbing contributed by other species such as hydrated lime or unoxidized sulfite in a forced oxidation system.

Run 951-2D. Because of the apparently high SO_2 removal at an inlet pH of 5.0, the pH was further reduced to 4.5. At this low pH adipic acid should not be fully effective and a drop in SO_2 removal should result. The SO_2 removal indeed dropped to 85 percent at a decreased inlet SO_2 concentration of 2150 ppm. The combination of low pH and high air stoichiometry of 2.05 atoms oxygen/mole SO_2 absorbed resulted in a good oxidation of 96 percent. The high air stoichiometry was the result of a drop in both SO_2 removal and inlet SO_2 concentration.

Run 951-2E. The purpose of Run 951-2E was to reconfirm the excellent SO_2 removal obtained at the low inlet pH of 5.0 during Run 951-2C. However, differences in system pressure drop (total pressure drop across the TCA minus mist eliminator pressure drop) between the two runs precluded exact replication.

In Run 951-2C, the system pressure drop was erratic and averaged about 11 inches H_2O. In Run 951-2E, the average system pressure drop was only 8.5 inches H_2O with the last two-thirds of the run operating below 8.0 inches H_2O. Somewhat higher than normal system pressure drop of 9 inches H_2O was also observed during Run 951-2B. Normal system pressure drop, under the operating conditions listed in Table 7-1, is usually 7 to 8 inches H_2O when the system is clean. The cause of the pressure drop variation was attributed to scaling and descaling of calcium sulfite.

At the lower system pressure drop during Run 951-2E, as compared to Run 951-2C, SO_2 removal dropped to 82 percent at 2400 ppm inlet SO_2 concentration.

It should be noted that an SO_2 removal of about 65 percent would be predicted under the same operating conditions as Run 951-2E but without forced oxidation and no adipic acid addition. Without adipic acid and with forced oxidation,

SO_2 removal should be lower than 65 percent.

Calcium Sulfite Scaling. As mentioned previously, the system pressure drop varied throughout this test series, which was most likely caused by scaling and descaling of calcium sulfite. Scaling appeared to begin during Run 951-2A at a high inlet pH of 7.0 and a high SO_2 make-per-pass of 14.4 m-moles/liter, and continued during Run 951-2B when the pH was reduced to 5.5 but at a rather high SO_2 make-per-pass of 12.8 m-moles/liter. Sulfite descaling began during Run 951-2C at an inlet pH of 5.0, and continued through Runs 951-2D (4.5 inlet pH) and 951-2E (5.0 inlet pH) at an SO_2 make-per-pass of about 10 m-moles/liter.

Several factors may have contributed to the sulfite scaling. Adipic acid buffered the pH drop across the scrubber and the higher outlet pH, especially at a high inlet pH of 7.0, caused sulfite supersaturation in the scrubber outlet liquor. High SO_2 make-per-pass, from the high SO_2 removal due to adipic acid addition, undoubtedly contributed to the sulfite supersaturation in the scrubber outlet liquor. Forced oxidation depleted the $CaSO_3$ solids which would otherwise be available for sulfite desupersaturation.

SUMMARY

The following is a summary of the test results:

- Within-scrubber-loop forced oxidation of sulfite to sulfate is feasible in a single-loop adipic acid-enhanced lime system. Under this condition calcium adipate becomes the major scrubbing species.
- At 1200 ppm adipic acid concentration and with forced oxidation, both good SO_2 removal and good oxidation greater than 90 percent were achieved under the operating conditions typical for the TCA. Optimum scrubber inlet pH appeared to be in the range of 5.0 to 5.5.
- Calcium sulfite supersaturation and the resultant scaling could occur under combinations of high pH, forced oxidation (low $CaSO_3$ solids concentration) and high SO_2 make-per-pass.

Section 8

LIMESTONE/ADIPIC ACID TESTING ON THE TCA WITH ONE TANK AND FORCED OXIDATION

This series of tests was originally intended to be an agitator study on the effects of various agitator speeds and air rates on the degree of sulfite oxidation. During the initial run, however, the system behaved as though the limestone was being blinded. The limestone feed rate rose sharply as the attempt was made to maintain the TCA inlet pH above 5.1. Limestone utilization ranged as low as 25 percent and averaged 38 percent. Concurrent with the drop in limestone utilization was a substantial increase in the apparent sulfite ion concentration (total oxidizable sulfite by amperometric titration) in the liquor entering the scrubber. The average inlet $SO_3^=$ concentration rose to 1510 ppm. Such concentrations are an order of magnitude above the normal sulfite values which average below 200 ppm.

Under such abnormal conditions, the plans for an agitator study were postponed. The agitator study was replaced by a series of tests designed to give additional information as to the cause of the abnormal system behavior.

The flow configuration for the adipic acid-enhanced limestone tests using one tank and forced oxidation is the same as that shown in Figure 7-1. Details of the oxidation tank arrangement are shown in Figure 7-2.

DISCUSSION OF TEST RESULTS

Table 8-1 lists the major test conditions and run-average results for all runs made in the test series. Major independent variables investigated were air flow rate (air stoichiometry) and limestone and adipic acid addition points, to observe their effects on system instability with respect to high liquor sulfite concentration and low limestone utilization (low pH). The effect of oxidation tank residence time was also investigated briefly. All runs were made with 1500 ppm adipic acid, 15 percent slurry solids concentration, and flue gas with high fly ash loading.

Effect of Air Stoichiometry. In the initial phase of Run 901-2A (see Part 1 for this run in Table 8-1) during which the abnormal conditions described above occurred (38 percent limestone utilization and 1510 ppm inlet liquor sulfite concentration), both the fresh limestone and adipic acid were fed to the oxidation tank as shown in Figure 7-1. Air was fed to the oxidation tank at a

Table 8-1

RESULTS OF TCA LIMESTONE/ADIPIC ACID TESTS WITH FORCED OXIDATION USING ONE TANK

	901-2A					901-2H	901-2I	901-2J		902-2A	902-2B
	1	2	3	4	5			1	2		
Major Test Conditions											
Fly ash loading	High					High	High	High	High	High	High
Adipic acid concentration, ppm	1500					1500	1500(2)	1500	1500	1500	1500
Gas rate, acfm @ 300°F	30,000					30,000	30,000	30,000	30,000	22,500	22,500
Liquor rate, gpm	1200					1200	1200	1200	1200	900	900
Percent solids recirculated	15					15	15	15	15	15	15
TCA inlet pH (controlled)	- (1)	- (1)	5.2	- (1)	5.2	5.2	5.4(3)	-	5.2	-	-
Limestone stoich. ratio (controlled)	1.2	1.2		1.2	-	-	-	1.2	-	1.2	1.2
Residence time, min.	4.1					4.1	4.1	4.1	4.1	5.5	5.5
Tank level, ft.	17					17	17	17	17	17	17
Limestone addition point(4)	D-204					G-201	G-201	D-204	D-204	D-204	D-204
Air rate to oxidizer, scfm	185	0	185	300	185	185	185	0	160(5)	110(5)	135(6)
Run-Average Results											
Start-of-run date	4/6/79	4/10/79	4/11/79	4/12/79	4/14/79	4/15/79	4/19/79	4/25/79	4/27/79	4/28/79	4/30/79
Onstream hours	91	23	28	45	21	94	92	63	33	51	39
Percent SO_2 removal	93.6	95.5	95.4	97.3	97.1	97.3	96.9	92.7	93.0	91.8	88.5
Inlet SO_2 concentration, ppm	2545	2635	2400	2530	2400	2275	2300	2460	2490	2255	2575
SO_2 make-per-pass, m-moles/l	12.6	13.1	12.1	13.2	12.6	11.5	11.8	12.0	11.9	11.2	11.8
Inlet liquor $SO_3^=$ concentration, ppm	1510	320	810	85	825	220	330	325	1055	675	1145
TCA inlet liquor percent gypsum sat'n	135	125	130	110	130	110	120	120	120	130	135
Slurry percent solids	14.6	13.7	14.6	13.6	15.2	15.2	15.0	13.9	14.9	14.8	14.7
Percent sulfite oxidation	97.5	17.5	96.0	99.3	97.5	99.0	98.5	24.0	94.0	98.6	97.5
Percent limestone utilization	38	84	55	88	41	65	47	86	45	50	56
Inlet pH	5.0	5.3	5.2	5.3	5.4	5.3	5.3	5.4	5.1	5.2	5.1
Adipic acid concentration, ppm	1530	1482	1395	1495	1390	1575	1530	1525	1485	1550	1450
Centrifuge cake solids, wt %	77	60	78	73	78	77	78	63	75	78	77
Air stoichiometry, atoms O/mole SO_2 abs.	1.60	0	1.70	2.45	1.65	1.75	1.70	0	1.45	1.45	1.60
Mist eliminator restriction, %	-	-	-	-	-	-	0	-	-	-	0

Notes:
(1) pH underride 5.2 changed to 4.9 early in run.
(2) Adipic acid added to process water hold tank.
(3) Changed during run to 5.2.
(4) D-204 is the oxidation tank while G-210 refers to the suction of the slurry recirculation pump.
(5) Air controlled to maintain a 1.45 air stoich. with a 1.55 max.
(6) Air controlled to maintain a 1.60 air stoichiometry.

rate of 185 scfm which gave an air stoichiometric ratio of 1.6 atoms oxygen/mole SO_2 absorbed. Average adipic acid concentration was 1530 ppm.

In the second phase of Run 901-2A the air was turned off to find out if the unusual system behavior would continue or if the system would return to normal. The TCA had been operated successfully many times before without oxidation using the same configuration (Reference 1). Within hours after the air had been stopped, the sulfite levels dropped to a near normal level of 320 ppm and limestone utilization increased to 84 percent. The SO_2 removal remained high at better than 95 percent with an inlet SO_2 concentration at 2635 ppm. The sulfite oxidation in solids dropped to 18 percent, normal for a run without forced oxidation.

The air was again introduced to the oxidizer at a rate of 185 scfm during the third phase of Run 901-2A. The purpose was to see if the unusual behavior experienced in phase 1 was a transient phenomenon or could be repeated. The answer came within a few hours as the inlet pH level fell and limestone addition rate had to be increased to maintain the inlet pH at 5.2. Liquor sulfite level also began to climb. When the system finally leveled out, the limestone utilization had dropped to 55 percent and liquor sulfite level had reached an average at 810 ppm. As with the first two phases, the SO_2 removal remained unaffected at greater than 95 percent with 2400 ppm SO_2 in the flue gas. The oxidation had returned to near its earlier value of 96 percent.

In the fourth phase of Run 901-2A, the air was increased to 300 scfm to see what effect the increased oxygen would have on the high liquor sulfite level. The resulting air stoichiometric ratio was 2.45 atoms oxygen/mole SO_2 absorbed. It was hypothesized that adipic acid or its degraded products might have formed complexes with sulfite in the liquor which were more difficult to oxidize. This hypothesis, however, could not be substantiated by the ion chromatograph and infrared analysis of the liquor species. Part 2 of Run 901-2J was made to provide the high sulfite samples for these analyses (Part 1 was made without air due to an air compressor failure).

Once again the results were quick to come. Within a few hours of air rate increase in Part 4 of Run 901-2A, the sulfite level dropped to only 85 ppm and limestone utilization rose to 88 percent. In addition, SO_2 removal increased to better than 97 percent despite a slight rise in inlet SO_2 concentration to 2530 ppm. As would be expected the oxidation reached a high of 99.3 percent.

To further document the unusual conditions experienced in phases 1 and 3 of Run 901-2A, and to provide additional samples of the liquor, the air rate was once more returned to the original rate of 185 scfm during the fifth phase of Run 901-2A. Again the system responded with an increased liquor sulfite level of 825 ppm and more limestone was again required to maintain pH control. Limestone utilization dropped to 41 percent. SO_2 removal and oxidation remained high, both at 97 percent.

Effect of Limestone Addition Point. Based on the results of Run 901-2A it was apparent that the high sulfite/low limestone utilization phenomenon was real and reproducible. A close examination of the IERL-RTP pilot plant test results (Reference 2) with limestone/adipic acid and with single-loop forced oxidation indicated that the high sulfite/low limestone utilization phenomenon observed at Shawnee did not exist in the pilot plant where limestone was added to a separate hold tank downstream of the oxidation tank.

In Run 901-2H* the limestone addition point was moved to the suction of the recirculated slurry pump. This removed the limestone from the oxidation tank but did not provide the additional residence time found in the IERL-RTP pilot plant tests utilizing a second hold tank. The resulting configuration was somewhere between that used in 901-2A and that used at the IERL-RTP pilot plant. The results reflected the compromise. While the average liquor sulfite level was reduced to 220 ppm, the limestone utilization improved but remained low at 65 percent. The modification did not affect the SO_2 removal or the oxidation which remained at 97 percent and 99 percent, respectively.

Effect of Adipic Acid Addition Point. A second difference between the IERL-RTP equipment configuration (Reference 2) and that used at Shawnee was that the addition of adipic acid was also outside the oxidation tank. At the IERL-RTP pilot plant the acid was added in a small receiving tank between the scrubber and the oxidation tank. Since one theory which had been advanced was that the sulfite might be complexing with the adipic acid or its degraded products, relocation of the adipic acid feed might have an effect on the test results. In Run 901-2I the adipic acid was added to the process water hold tank which received the clarifier overflow and centrate. The results showed no apparent effect over that experienced in Run 901-2H. The sulfite level increased slightly to 330 ppm while limestone utilization dropped to 47 percent. The SO_2 removal was 97 percent and oxidation was 98.5 percent.

Effect of Oxidation Tank Residence Time. In Part 4 of Run 901-2A, it had been demonstrated that sufficient air could oxidize the liquor sulfite that was being generated. In Runs 902-2A and 902-2B, both the flue gas and slurry flow rates were reduced to 22,500 acfm and 900 gpm, respectively, to increase the oxidation tank residence time (with respect to both slurry bleed rate and recirculation rate) in the hope that this would facilitate the oxidation of liquor sulfite. Unfortunately, the magnitude of change was small and the higher residence time of 5.5 minutes (with respect to 900 gpm recirculated slurry rate) was apparently insufficient to cause a significant change in the test results. The SO_2 removal dropped to about 90 percent for the two runs at the reduced slurry flow rate of 900 gpm.

Limestone Blinding and $CaSO_3$ Supersaturation. The phenomenon of high liquor sulfite/low limestone utilization can best be explained by referring to the discussions presented in Sections 5 and 7 on the same subject. High levels of solids oxidation, as in the within-loop-oxidation of a single-loop system or second loop of a two-loop system with recycle of oxidized slurry from the first loop (oxidation loop), leave insufficient $CaSO_3$ solid crystals for sulfite desupersaturation. If a particular system configuration, as in the second loop of a two-loop system, requires $CaSO_3$ precipitation in the scrubber, excessive $CaSO_3$ supersaturation would result. Even if $CaSO_3$ precipitation is not required, as in the present test series with a single loop and forced oxidation, high levels of dissolved sulfite, and therefore supersaturation, can result from insufficient oxidation intensity.

---

*Runs 901-2B through 901-2G were originally part of the scheduled agitator study and were not made because of the change of the test objective in this test series.

Excessive $CaSO_3$ supersaturation results in $CaSO_3$ scaling and limestone blinding. The latter subsequently causes low pH and poor limestone utilization.

In a single-loop system with forced oxidation, however, the effect of $CaSO_3$ supersaturation can be avoided by operating the oxidizer with excess air (as in Part 4 of Run 901-2A). Single-loop operation without forced oxidation should give sufficient $CaSO_3$ solids for sulfite desupersaturation (as in Run 901-2A when no air was used).

Adipic acid addition intensifies the $CaSO_3$ supersaturation because of the higher SO_2 make-per-pass it produces and the decreased pH drop across the scrubber (due to the buffering effect of adipic acid).

SUMMARY

The following is a summary of the results of testing on the TCA with limestone/adipic acid with forced oxidation using one tank:

- Calcium sulfite supersaturation and blinding of limestone, resulting in low pH and/or poor limestone utilization, could result from insufficient oxidation intensity in a single-loop forced oxidation system. Air stoichiometry necessary to avoid this situation appeared to be greater than about 1.7 atoms oxygen/mole SO_2 absorbed under the conditions tested (1500 ppm adipic acid). Adipic acid addition probably contributed to the calcium sulfite supersaturation due to the pH buffering effect (reduced pH drop across the TCA) and higher SO_2 make-per-pass.

- Despite the high liquor sulfite/low limestone utilization phenomenon, SO_2 removal greater than 92 percent was achieved at 1500 ppm adipic acid concentration under typical TCA operating conditions. Sulfite oxidation greater than 95 percent was obtained with air stoichiometries as low as 1.45 atoms oxygen/mole SO_2 absorbed.

Section 9

LIMESTONE/ADIPIC ACID TESTING ON THE TCA WITH TWO TANKS AND FORCED OXIDATION

Adipic acid-enhanced limestone test results from the TCA system with forced oxidation using a single tank (Section 8) indicated that high liquor sulfite/ low limestone utilization could result from insufficient oxidation intensity. Limitation in the size of the oxidation tank prevented testing of the effect of higher oxidation tank residence times (Runs 902-2A and 902-2B, Table 8-1).

As mentioned in Section 8, limestone/adipic acid tests at the IERL-RTP pilot plant (Reference 2) with forced oxidation using two tanks (oxidation tank followed by a second hold tank where limestone was added) did not produce the high liquor sulfite/low limestone utilization phenomenon observed at Shawnee. Operation with two tanks in series with limestone added to the second tank has several advantages over the single-tank operation:

- Low pH (scrubber-effluent pH) in the first tank (oxidation tank) promotes sulfite oxidation.
- The possibility of limestone blinding by calcium sulfite is decreased because limestone is added after the oxidation tank.
- Limestone utilization is increased with two tanks in series which approximate a plug flow reactor.
- Extra residence time for calcium sulfate crystallization is provided by the second tank.

SYSTEM DESCRIPTION

Figure 9-1 shows a flow diagram for the adipic acid-enhanced limestone scrubbing in the single-loop TCA system with forced oxidation using two tanks. Adipic acid was added as a dry powder to the oxidation tank (the first tank) while limestone was added to the effluent hold tank (the second tank). The detailed arrangement of the oxidation tank is shown in Figure 7-2. The external downcomer of the oxidation tank was used in all runs.

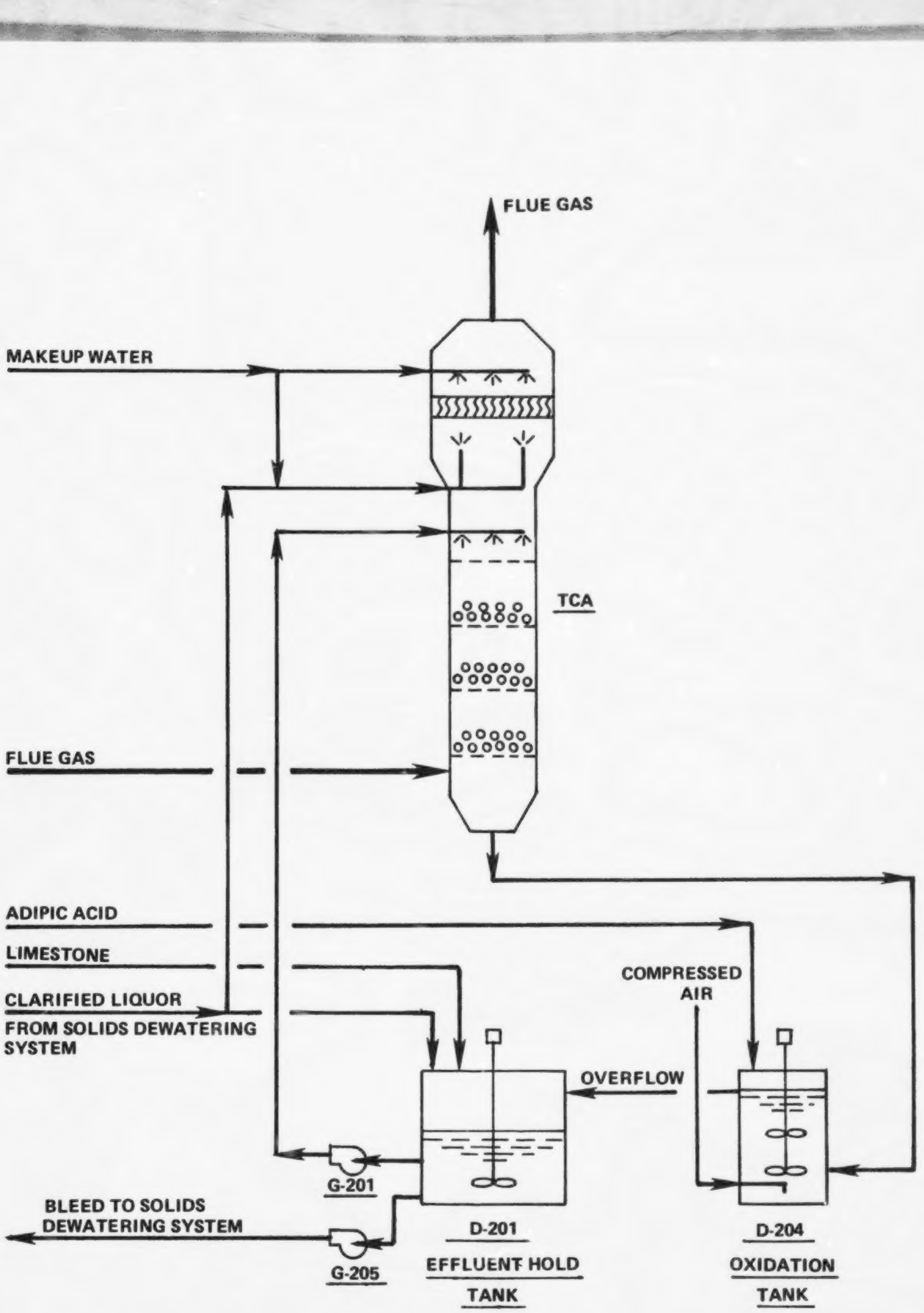

FLUE GAS
MAKEUP WATER
TCA
FLUE GAS
ADIPIC ACID
LIMESTONE
CLARIFIED LIQUOR
FROM SOLIDS DEWATERING
SYSTEM
COMPRESSED
AIR
OVERFLOW
G-201
BLEED TO SOLIDS
DEWATERING SYSTEM
G-205
D-201
EFFLUENT HOLD
TANK
D-204
OXIDATION
TANK

Figure 9-1. FLOW DIAGRAM FOR ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING IN THE TCA SYSTEM WITH TWO TANKS AND FORCED OXIDATION

DISCUSSIONS OF TEST RESULTS

Two runs (Runs 903-2A and 903-2B) were made with 1500 ppm adipic acid in the first half of May 1979 before a series of limestone/adipic acid factorial tests (Section 10). Four additional runs (Run 906-2A through 906-2D) were conducted in mid-October 1979 to cover a wider range of adipic acid concentration, including an adipic acid depletion test from 2000 ppm to 800 ppm and one from 800 ppm to 400 ppm. Table 9-1 summarizes the major test conditions and run-average results.

Effect of Two-tank vs. One-tank Operation. Run 903-2A (Table 9-1) was made under the same operating conditions as Run 901-2A (Table 8-1) except for the number of tanks used. Relocation of the limestone feed to the second tank in Run 903-2A proved to be successful in solving the high liquor sulfite/low limestone utilization problem experienced in the single-tank mode. A comparison of the results from Part 1 of Run 901-2A and Run 903-2A clearly shows the advantages of two-tank operation:

	Run 901-2A Part 1	Run 903-2A
Number of tanks	1	2
Percent SO_2 removal	93.6	93.8
Inlet SO_2 conc., ppm	2545	2680
SO_2 make-per-pass, m-moles/l	12.6	13.2
Inlet liquor $SO_3^=$ conc., ppm	1510	55
Inlet liquor percent gypsum sat'n	135	95
Percent sulfite oxidation	97.5	99.1
Percent limestone utilization	38	90
Inlet pH	5.0	5.3
Adipic acid conc., ppm	1530	1550
Air stoich., atoms O/mole SO_2 abs.	1.60	1.50

With slightly lower air stoichiometry of 1.50 atoms oxygen/mole SO_2 absorbed, two-tank operation gave only 55 ppm of sulfite in the TCA inlet liquor compared to 1510 ppm for the one-tank mode (Run 901-2A). Limestone utilization was 90 percent with a TCA inlet pH of 5.3 for the two-tank mode, compared to 38 percent utilization and 5.0 inlet pH for the one-tank scheme. Higher sulfite oxidation and lower gypsum saturation were also observed for Run 903-2A.

For all runs made with two tanks, average inlet liquor sulfite never exceeded 100 ppm and limestone utilizations were all above about 80 percent. Higher pH's and lower gypsum saturations in the TCA inlet liquor were also observed. Sulfite oxidations were all greater than 99 percent.

Effect of Flue Gas Source. Based on the success of Run 903-2A, the two-tank configuration was chosen for the limestone/adipic acid forced oxidation factorial series (Section 10) to be operated on flue gas from Boiler No. 9 during the 9-week maintenance outage of Boiler No. 10. In order to check the effect of the gas from Boiler No. 9, Run 903-2B was made under the same operating conditions as Run 903-2A except for the change of flue gas. The

Table 9-1

RESULTS OF TCA LIMESTONE/ADIPIC ACID TESTS WITH FORCED OXIDATION USING TWO TANKS

Major Test Conditions	903-2A	903-2B	906-2A(3)	906-2B	906-2C	906-2D(3)
Fly Ash loading	High	High	High	High	High	High
Adipic acid concentration, ppm	1500	1500	2000→800	800	800	800→400
Gas rate, acfm @ 300°F	30,000	30,000(1)	25,000	25,000	25,000	25,000
Liquor rate, gpm	1200	1200	1200	1200	1200	1200
Percent solids recirculated	15	15	15	15	15	15
Limestone stoich. ratio (controlled)	1.2	1.2	1.2	1.2	1.2	1.2
Res. time (min)/Tk level (ft):						
Oxid Tk (D-204)	4.3/18	4.3/18	4.3/18	4.3/18	4.3/18	4.3/18
EHT (D-201)	12/6.1	12/6.1	12/6.1	12/6.1	12/6.1	12/6.1
Air rate to oxidizer, scfm	185	185	185	185	185	185
Run-Average Results						
Start-of-Run Date	5/2/79	5/9/79	10/11/79	10/17/79	10/18/79	10/19/79
Onstream hours	51	88	158	24	24	48
Percent SO_2 removal	93.8	96.2	98.1→94.8	95.3	96.5	96.9→93.6
Inlet SO_2 concentration, ppm	2680	2370	1940	2310	2220	2050
SO_2 make-per-pass, m-moles/l	13.2	12.1	8.3	9.7	9.5	8.6
Inlet liquor $SO_3^=$ concentration, ppm	55	70	40	95	25	30
Inlet liquor percent gypsum sat'n	95	100	95	90	95	90
Slurry percent solids	14.4	15.2	15.0	15.1	14.8	14.8
Percent sulfite oxidation	99.1	99.5	99.2	99.3	99.5	99.6
Percent limestone utilization	90	86	82	88	78	81
Inlet pH	5.30	5.42	5.72	5.86	6.10	5.73
Adipic acid concentration, ppm	1550	1530	2115→830	770	850	860→600
Filter cake solids, wt %	78(2)	88(4)	86	(5)	88	87
Air stoich., atoms O/mole SO_2 abs.	1.50	1.65	2.40	2.05	2.10	2.30
Mist eliminator restriction, %	0	-	-	-	-	-

(1) Flue gas from Unit 9.
(2) Centrifuge cake solids.
(3) Adipic acid depletion tests.
(4) Horizontal belt filter cake solids.
(5) Open-loop operation.

results for the two runs were consistent and no significant differences were discernible.

Effect of Adipic Acid Concentration on SO_2 Removal. In Runs 906-2A through 906-2D, adipic acid concentration was decreased from 2000 ppm to 400 ppm to observe its effect on SO_2 removal.

Adipic acid was depleted from 2000 ppm to 800 ppm during Run 906-2A by reducing the acid addition rate under closed-liquor-loop condition. It was then maintained at 800 ppm under closed-liquor-loop condition during Run 906-2C. During Run 906-2D, adipic acid concentration was to be further depleted from 800 ppm to 400 ppm, again under closed-loop condition and at reduced addition rate. Final concentration, however, was 600 ppm, instead of the planned 400 ppm, because the TCA system had to be transferred to TVA for the demonstration of the DOWA basic aluminum oxide scrubbing process.

During this series of tests (Runs 906-2A, 906-2C, and 906-2D), actual adipic acid concentration decreased from 2115 to 600 ppm while the SO_2 removal dropped from 98.1 to 93.6 percent, demonstrating good SO_2 removal with adipic acid concentration as low as 600 ppm. Average inlet SO_2 concentration, however, was low during this series, ranging from 1940 to 2220 ppm.

Effect of Open-Liquor-Loop Operation. Run 906-2B was conducted with open-liquor-loop operation. The scrubber bleed slurry (15 percent solids) was discharged directly from the system, bypassing the clarifier and the rotary drum vacuum filter with no clarified liquor or filtrate returned to the scrubber loop. This mode of operation requires a much higher adipic acid addition rate, and since no degraded acid is returned by the clarified liquor or filtrate, the adipic acid existing in the scrubber loop should be relatively fresh and not degraded. Therefore, the effect of the "freshness" of adipic acid on SO_2 removal can be compared between the "fresh" acid in the open-loop operation (Run 906-2B) and the "degraded" acid in the closed-loop operation (Run 906-2C).

Little or no effect on the SO_2 removal was observed. The SO_2 removal was 96.5 percent at 2220 ppm inlet SO_2 concentration for Run 906-2C with closed-loop operation, compared to 95.3 percent removal at 2310 ppm inlet SO_2 concentration for Run 906-2B with open-loop mode. This observation is consistent with that made during venturi/spray tower limestone/adipic acid factorial tests (Section 4).

SUMMARY

The following is a summary of the results of limestone/adipic acid tests with forced oxidation using two tanks:

- Two-tank operation is superior to one-tank operation. With the two-tank mode, the high liquor sulfite/low limestone utilization problem experienced with one-tank operation was eliminated. Furthermore, higher sulfite oxidation, higher scrubber inlet pH, and lower gypsum saturation were achieved with the two-tank scheme.

- No differences in the test results were discernible when flue gas from Boiler No. 9 or Boiler No. 10 was used.

- Good SO_2 removal of 93.6 percent was achieved with adipic acid concentration as low as 600 ppm at 2050 ppm inlet SO_2 concentration, liquid-to-gas ratio of 60 gal/mcf, and 81 percent limestone utilization.

- At 800 ppm concentration, "freshness" (fresh vs. degraded) of adipic acid did not show an effect on SO_2 removal.

Section 10

FACTORIAL TESTS ON THE TCA SYSTEM WITH LIMESTONE/ADIPIC ACID SLURRY

A series of 64 limestone/adipic acid factorial tests (Runs TAA001 through TAA064) were conducted on the TCA system. All but six of these runs were made with forced oxidation using two tanks. The flow configuration used for all the tests is shown in Figure 9-1. The detailed arrangement of the oxidation tank is shown in Figure 7-2. The external downcomer of the oxidation tank was used.

Major variables investigated included adipic acid concentration, gas and slurry flow rates (liquid-to-gas ratio), total bed height (zero or 15 inches), and scrubber inlet pH. The test results are listed in Table 10-1.

The 64 tests were divided into two groups:

- Testing at a controlled limestone stoichiometry of 1.2, giving a high scrubber inlet pH of 5.6 to 6.1 (Runs TAA001 through TAA052).
- Testing at controlled lower scrubber inlet pH of 4.6 to 5.6 (Runs TAA053 through TAA064).

Operating conditions common to all tests were:

Fly ash loading: High (3-6 grains/dry scf)
Slurry solids concentration: 15 percent
Oxidation tank level (7 ft diameter): 18 ft
Effluent hold tank level (20 ft diameter): 6.2 ft
Air flow to oxidizer: 220 scfm (for runs with forced oxidation)
Oxidation tank agitator speed: 54 rpm

LIMESTONE/ADIPIC ACID SCRUBBING AT HIGH pH

Equation 10-1 below for the prediction of SO_2 removal has been fitted to the 52 TCA limestone/adipic acid factorial tests (Runs TAA001 through TAA052) made at a controlled limestone stoichiometry of 1.2 moles Ca/mole SO_2 absorbed. All but six of these runs were with forced oxidation. Eight runs (Runs TAA045 through TAA052) were base case runs without adipic acid additive.

Table 10-1

TCA LIMESTONE/ADIPIC ACID FACTORIAL TEST RESULTS

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)	L/G (gal/mcf)	Press. Drop (in. H_2O)	TCA Inlet Liquor pH		TCA Outlet Liq. pH	Ave. AD Conc. (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration, (ppm)		Percent SO_2 Removal		Air Rate (scfm)	Sulfite Oxid. (%)	Replicate	Spheres (inches)
					Range	Ave.				Range	Ave	Range	Ave.				
TAA001	25,000	898	44	6.3	5.66-5.76	5.71	5.0	648	1997	2000-2060	2014	89-91	89.9	220	100.0	A	15
TAA002	20,000	1184	73	5.9	5.61-5.74	5.66	5.1	651	2086	1860-2080	1919	96-97	96.4	220	99.6		15
TAA003	20,000	598	37	4.5	5.53-5.82	5.69	4.8	652	2130	1940-2320	2109	80-82	81.3	220	99.6		15
TAA004	30,000	904	37	7.5	5.65-5.67	5.66	4.8	618	2308	2120-2320	2219	88-91	89.4	220	99.6		15
TAA005	20,000	898	55	5.3	5.84-6.06	5.92	5.2	624	1864	2000-2300	2160	91-92	91.5	220	99.6		15
TAA006	25,000	906	45	6.4	5.66-5.88	5.80	4.8	635	1953	1960-2140	2171	90-92	90.8	220	100.0	A	15
TAA007	30,000	1186	49	8.9	5.67-5.75	5.72	5.0	595	2108	2120-2280	2169	93-95	93.9	220	99.5		15
TAA008	30,000	603	25	6.3	5.75-5.82	5.79	4.9	591	2485	2200-2440	2285	77-80	78.3	220	99.5		15
TAA009	25,000	1181	58	7.2	5.71-5.82	5.75	5.0	646	2418	2320-2520	2400	93-95	94.2	220	99.6		15
TAA010	25,000	597	29	2.0	5.97-6.02	6.00	5.0	644	2308	2400-2440	2420	59-61	59.5	220	99.1		0
TAA011	25,000	888	44	2.4	5.94-6.07	6.00	5.2	644	2374	2120-2200	2184	70-72	71.2	220	99.8		0
TAA012	25,000	1187	59	2.8	5.96-6.00	5.98	5.3	620	2352	2300-2400	2362	77-80	78.1	220	100.0		0
TAA013	25,000	1197	59	7.0	6.01-6.02	6.01	5.4	667	2219	2400-2440	2429	93-94	93.9	0	43.6		15
TAA014	25,000	590	29	5.2	6.11-6.14	6.12	5.2	529	1997	2600-2640	2631	77-80	78.4	0	36.5		15
TAA015	25,000	594	29	5.2	5.90-6.13	6.04	5.2	1389	1697	2320-2400	2340	87-88	87.8	0	30.3		15
TAA016	25,000	1176	58	7.0	5.89-6.07	5.97	5.3	1309	1598	2320-2520	2432	97-99	97.3	0	25.4		15
TAA017	25,000	894	44	2.3	5.93-6.26	6.05	5.3	1176	1790	2320-2780	2523	78-80	78.6	220	97.3	C	0
TAA018	25,000	587	29	2.1	6.07-6.23	6.13	5.1	1415	2130	2400-2460	2421	70-72	71.1	220	98.1		0
TAA019	25,000	1190	59	2.5	5.99-6.06	6.03	5.4	1423	2352	2440-2520	2474	86-87	86.2	220	99.1		0
TAA020	25,000	883	44	2.5	5.98-6.10	6.04	5.3	1316	2041	2680-2800	2750	78	78.0	220	99.6	C	0
TAA021	25,000	901	44	6.3	5.87-5.93	5.89	5.1	1273	2375	2280-2540	2398	92-94	93.5	220	99.3	D	15
TAA022	20,000	895	55	5.1	5.87-6.10	6.01	5.1	1144	1775	2720-2880	2804	93-95	94.0	220	99.9		15

Table 10-1 (Continued)

TCA LIMESTONE/ADIPIC ACID FACTORIAL TEST RESULTS

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)	L/G (gal/mcf)	Press. Drop (in. H_2O)	TCA Inlet Liquor pH		TCA Outlet Liq. pH	Ave. AD$^=$ Conc. (ppm)	Ave. Cl$^-$ Conc. (ppm)	Inlet SO_2 Concentration, (ppm)		Percent SO_2 Removal		Air Rate (scfm)	Sulfite Oxid. (%)	Replicate	Spheres (inches)
					Range	Ave.				Range	Ave.	Range	Ave.				
TAA023	30,000	898	37	7.4	5.79-5.85	5.82	5.0	1384	1909	2460-2600	2546	93-94	93.9	220	99.8		15
TAA024	25,000	580	29	5.1	5.92-6.01	5.95	5.0	1287	1997	2600-2800	2651	84-86	85.6	220	99.7	E	15
TAA025	25,000	885	44	6.1	5.92-6.04	5.97	5.0	1287	2086	2160-2560	2378	94-96	95.2	220	99.5	D	15
TAA026	25,000	1192	59	7.1	5.82-5.91	5.87	5.2	1335	2041	1840-2080	1942	97-98	97.3	220	99.8		15
TAA027	25,000	604	29	5.5	5.87-6.01	5.94	5.0	1247	1930	2400-2680	2461	85-87	86.3	220	99.2	E	15
TAA028	25,000	893	44	6.1	5.76-5.89	5.85	5.1	2122	1331	2340-2720	2524	96-97	96.2	220	99.7	F	15
TAA029	25,000	601	29	5.2	5.97-6.06	6.01	5.0	1911	1287	2360-2400	2373	90-91	90.6	220	99.6	G	15
TAA030	30,000	905	37	7.6	5.49-5.64	5.56	5.0	1945	1598	2740-2960	2861	93-95	93.9	220	99.6		15
TAA031	20,000	901	55	5.0	5.86-5.91	5.88	5.1	2088	1442	2440-2480	2454	97-98	97.2	220	99.8		15
TAA032	25,000	900	44	6.1	5.79-5.82	5.80	5.1	1878	1775	2320-2520	2484	95-96	95.4	220	100.0	F	15
TAA033	25,000	1191	59	7.1	5.73-5.85	5.79	5.2	2116	1753	2360-2520	2443	97-98	97.9	220	99.9		15
TAA034	25,000	589	29	5.2	5.79-5.86	5.82	5.0	1916	1686	2280-2360	2329	90-91	90.4	220	100.0	G	15
TAA035	25,000	1203	59	2.7	5.64-5.90	5.76	5.2	2065	1775	2160-2540	2389	86-89	88.1	220	100.0		0
TAA036	30,000	895	37	3.3	5.63-5.79	5.71	5.0	2174	1819	2400-2440	2417	82-87	83.9	220	99.2		0
TAA037	30,000	602	25	2.6	5.69-5.83	5.77	4.9	2099	1953	2300-2520	2404	72-75	72.7	220	99.5		0
TAA038	20,000	898	55	1.8	5.64-5.75	5.70	5.0	2254	1775	2320-2560	2407	86-88	86.4	220	99.7	H	0
TAA039	20,000	596	37	1.4	5.70-5.77	5.74	4.9	2025	1864	2480-2600	2540	77-78	77.4	220	100.0		0
TAA040	30,000	1191	49	3.6	5.62-5.72	5.68	5.1	2245	1864	2240-2400	2287	87-88	87.2	220	99.4		0
TAA041	20,000	896	55	1.7	5.73-5.76	5.75	5.0	1981	1775	2560-2680	2625	83-86	83.8	220	99.2	H	0
TAA042	25,000	903	44	2.3	5.60-5.75	5.68	5.0	2284	1886	2320-2720	2487	80-85	81.8	220	99.7		0
TAA043	25,000	597	29	5.3	5.90-5.97	5.94	5.2	1955	1893	2320-2640	2472	89-92	90.3	0	30.4		15
TAA044	25,000	1191	59	7.1	5.85-5.90	5.88	5.3	2128	1953	2240-2280	2253	97-98	97.9	0	33.7		15(1)

(1) Actual measured sphere depth at the end of TAA-044 was 12-1/2 inches.

Table 10-1 (Continued)

TCA LIMESTONE/ADIPIC ACID FACTORIAL TEST RESULTS

Run No.	Gas Rate (acfm)	Liquor Rate (gpm)	L/G (gal/mcf)	Press. Drop (in. H_2O)	TCA Inlet Liquor pH		TCA Outlet Liq. pH	Ave. $AD^=$ Conc. (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration, (ppm)		Percent SO_2 Removal		Air Rate (scfm)	Sulfite Oxid. (%)	Replicate	Spheres (inches)
					Range	Ave.				Range	Ave.	Range	Ave.				
TAA045	25,000	900	44	2.3	5.94-6.20	6.08	5.2	0	3432	2400-2560	2452	53-58	55.4	220	99.8	I	0
TAA046	25,000	1193	60	2.7	5.79-5.89	5.82	5.1	0	3586	2080-2360	2208	65-66	65.1	220	97.7		0
TAA047	25,000	592	30	1.7	6.04-6.10	6.07	5.2	0	3515	1680-1740	1769	46-48	47.1	220	99.4		0
TAA048	25,000	887	44	2.3	6.18	6.18	5.3	0	3248	2000-2200	2094	57-59	57.9	220	99.6	I	0
TAA049	25,000	902	44	6.4	5.92-6.27	6.06	5.3	0	2568	2200-2280	2262	71-76	73.7	220	99.5	J	15
TAA050	25,000	1178	59	7.5	6.05-6.13	6.09	5.0	0	3248	2080-2160	2110	83-84	83.6	220	99.5		15
TAA051	25,000	575	29	5.1	5.88-5.97	5.92	5.0	0	3728	2000-2100	2071	56-59	57.5	220	99.5		15
TAA052	25,000	904	44	6.5	5.81-6.05	5.96	5.0	0	3269	2320-2520	2429	73-76	74.9	220	99.0	J	15
TAA053	25,000	905	44	6.6	5.51-5.82	5.62	5.2	707	2101	2000-2400	2191	84-90	86.6	220	96.3		15
TAA054	25,000	906	44	6.6	5.07-5.16	5.11	4.7	761	4112	2120-2600	2452	64-83	72.1	220	98.7		15
TAA055	25,000	907	44	6.7	4.47-4.63	4.57	3.6	667	3550	2240-2360	2331	44-50	46.9	220	99.4		15
TAA056	25,000	906	44	6.7	5.22-5.41	5.29	4.7	1247	3432	2240-2440	2329	88-92	90.5	220	100.0		15
TAA057	25,000	905	44	6.6	5.45-5.52	5.50	5.0	1249	2870	2240-2520	2365	88-92	89.7	220	100.0		15
TAA058	25,000	904	44	6.5	5.10-5.41	5.25	4.6	1409	2958	2200-2480	2272	86-89	87.4	220	99.6	K	15
TAA059	25,000	898	44	6.8	4.91-5.09	5.01	4.2	1273	2973	2000-2880	2246	68-84	76.0	220	99.5		15
TAA060	25,000	900	44	6.9	4.43-4.67	4.52	3.6	1333	2556	2520-2960	2800	54-64	62.9	220	100.0		15
TAA061	25,000	905	44	6.9	5.06-5.32	5.23	4.7	1497	2645	2500-2640	2528	89-92	91.0	220	98.6	K	15
TAA062	25,000	903	44	6.5	5.21-5.45	5.37	4.9	2102	2944	2200-2280	2263	94-95	95.0	220	99.5		15
TAA063	25,000	899	44	6.5	4.95-5.13	5.03	4.6	2088	2781	1640-1920	1776	92-94	93.7	220	99.6		15
TAA064	25,000	901	44	6.7	4.46-4.60	4.57	3.9	2109	2603	1600-1760	1690	82-85	83.8	220	100.0		15

$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp[-0.0645\, L^{0.76} \exp(0.038H + 0.0177A^{0.5})] \quad (10\text{-}1)$$

where

L = slurry flow rate, gpm/ft^2
H = total static height of 1-7/8 inch diameter nitrile foam spheres, inches
A = adipic acid concentration in scrubber liquor, ppm

The ranges of operating variables covered by Runs TAA001 through TAA052, and therefore by Equation 10-1, are:

Slurry flow rate (L) = 19-37 gpm/ft^2 (600-1200 gpm)
Gas flow rate = 20,000-30,000 acfm @ 300°F
Total height of spheres (H) = 0 or 15 inches
Number of grids = 4
Adipic acid concentration in scrubber liquor (A) = 0-2300 ppm
Scrubber inlet pH = 5.6-6.1 (1.2 limestone stoichiometry)
Inlet gas SO_2 concentration = 1800-2800ppm
Sulfite oxidation = 95-100% for forced-oxidation runs, and 20-40% for runs without forced oxidation

Figure 10-1 shows measured values of SO_2 removal plotted against corresponding values predicted by Equation 10-1. For the 52 correlated runs, Equation 10-1 explains 98 percent of the variation in the data for SO_2 removal with a standard error of estimate of 1.8 percent SO_2 removal.

Figure 10-2 shows SO_2 removal as a function of adipic acid concentration and slurry flow rate for the TCA without spheres (grid tower). The lines in Figure 10-2 are predicted by Equation 10-1; the actual data points for no spheres are also shown in the figure. Figure 10-2 indicates that, for a slurry flow rate of 28 gpm/ft^2 (900 gpm for the Shawnee TCA), an increase in adipic acid concentration from zero to 1200 ppm increases SO_2 removal from 55 to 78 percent. Further increase of adipic acid concentration to 2000 ppm increases SO_2 removal to 85 percent. At an adipic acid concentration of 1200 ppm, an increase in slurry flow rate from 19 to 37 gpm/ft^2 increases SO_2 removal from 67 to 85 percent. Thus, for the grid tower configuration, either a doubling of the slurry flow rate or the addition of 1200 ppm of adipic acid cuts SO_2 emission by a factor of about two.

Figure 10-3 is a plot of SO_2 removal as a function of adipic acid concentration and slurry flow rate for the TCA with 15 inches of total static sphere height. Again, the lines in Figure 10-3 are predicted by Equation 10-1. For a slurry flow rate of 19 gpm/ft^2, adipic acid concentrations of zero, 600, and 2000 ppm give SO_2 removals of 60-65, 80, and 90 percent, respectively. For an adipic acid concentration of 600 ppm, doubling the slurry flow rate from 19 to 37 gpm/ft^2 increases SO_2 removal from 80 to 94 percent.

Figures 10-2 and 10-3 can be compared to obtain the effect of adding 15 inches of spheres on SO_2 removal in the TCA. At a slurry flow rate of 37 gpm/ft^2 and zero adipic acid concentration, the addition of spheres increases SO_2 removal from 63 to 83 percent, corresponding to a decrease in SO_2 emission

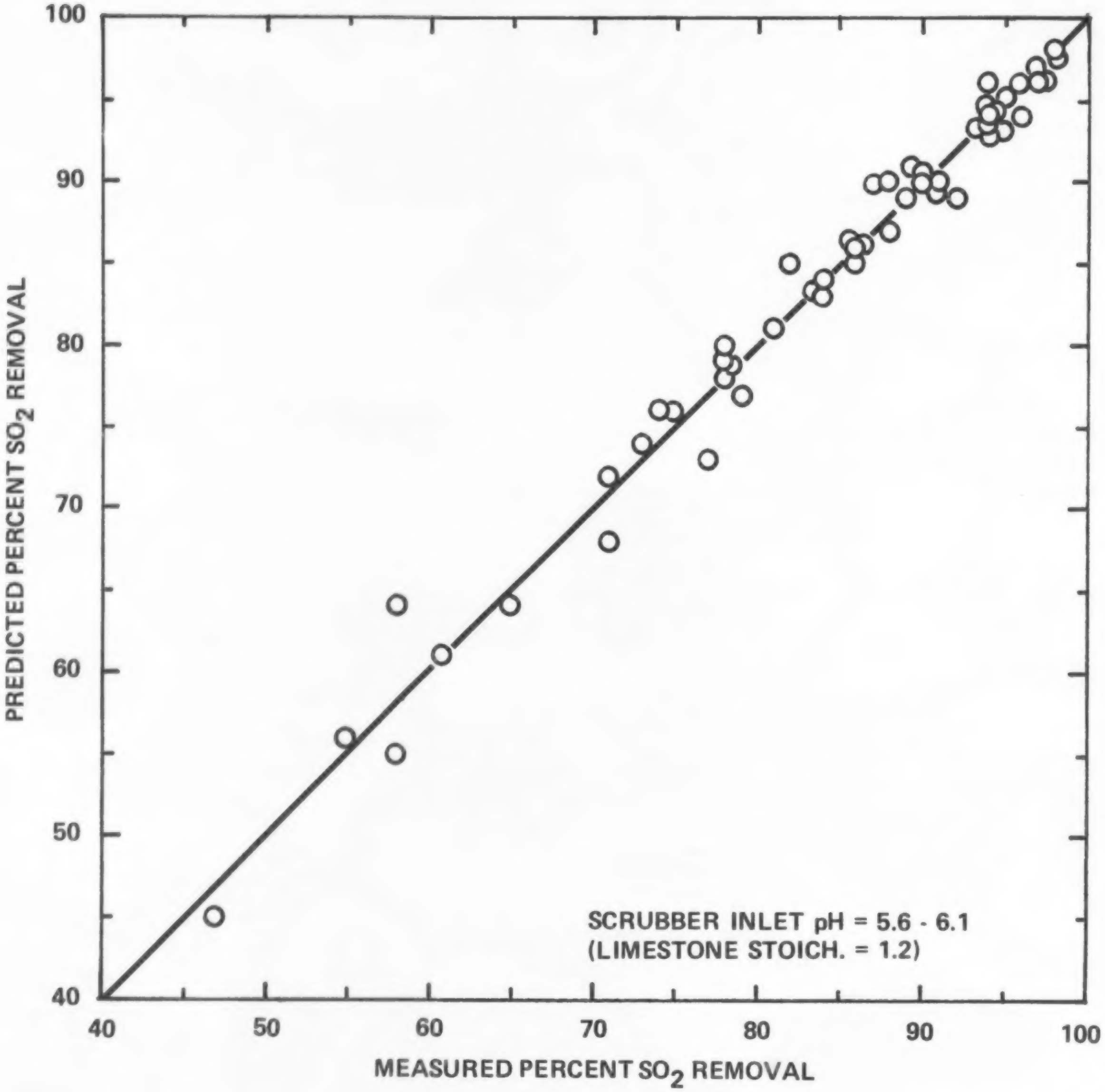


FIGURE 10-1

MEASURED SO_2 REMOVAL vs PREDICTED SO_2 REMOVAL BY EQUATION 10-1

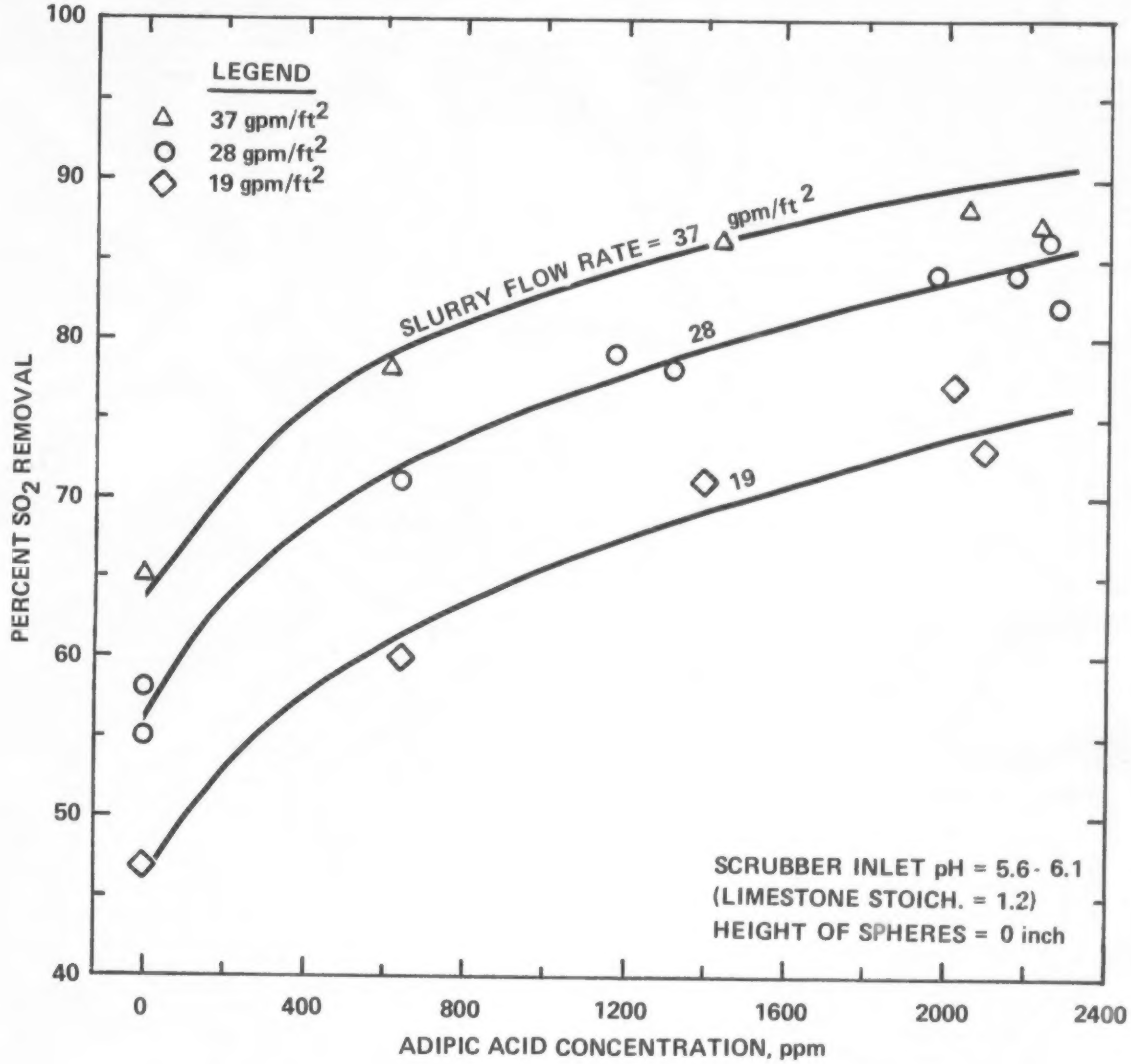


FIGURE 10-2

EFFECT OF ADIPIC ACID CONCENTRATION AND SLURRY FLOW RATE ON SO_2 REMOVAL IN THE TCA WITH FOUR GRIDS AND WITHOUT SPHERES

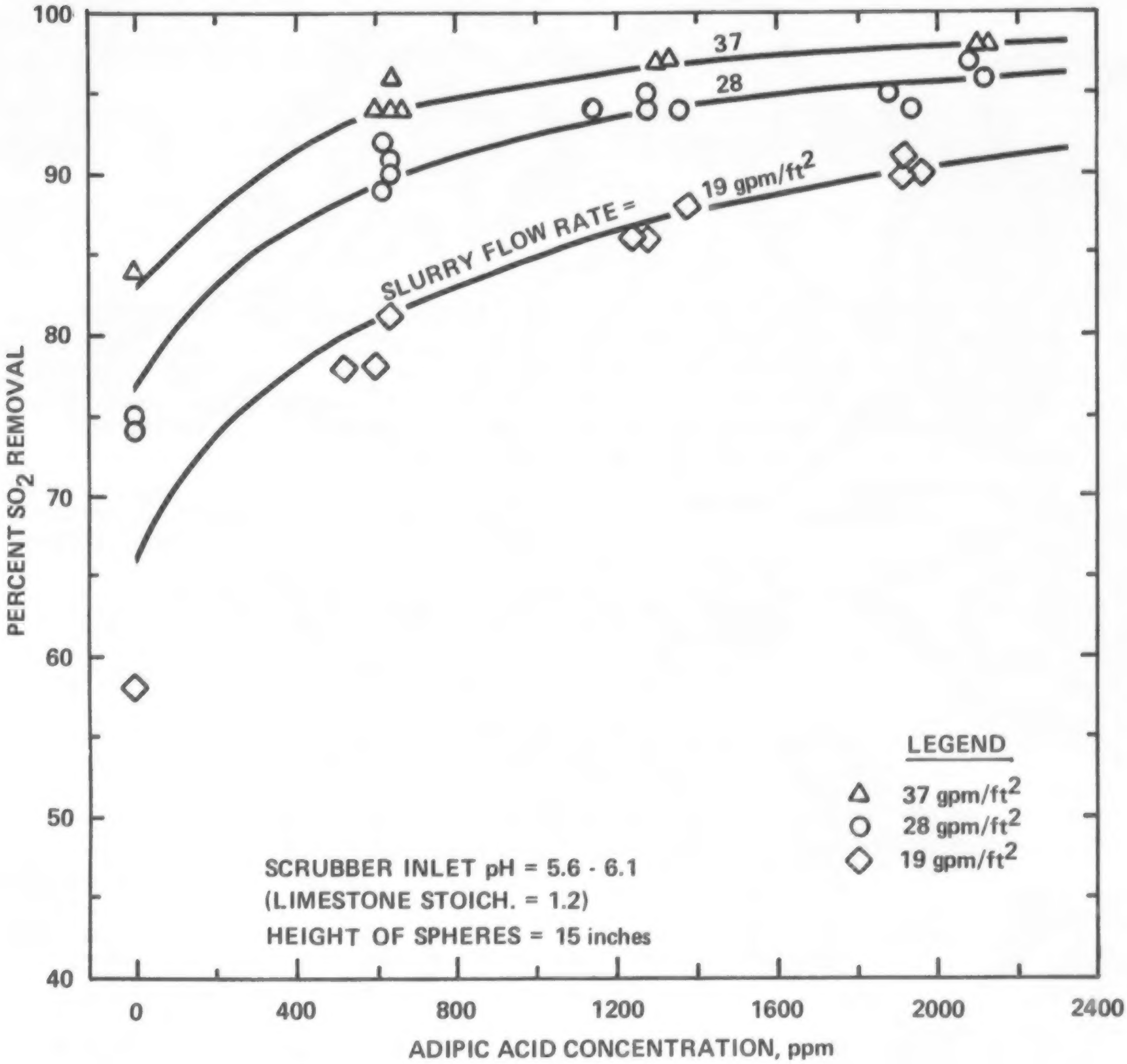


FIGURE 10-3

EFFECT OF ADIPIC ACID CONCENTRATION AND SLURRY FLOW RATE ON SO_2 REMOVAL IN THE TCA WITH FOUR GRIDS AND 15 INCHES OF SPHERES by a factor of two. At a slurry flow rate of 19 gpm/ft^2 with 2000 ppm adipic acid, 15 inches of spheres increase SO_2 removal from 74 to 90 percent.

LIMESTONE/ADIPIC ACID SCRUBBING AT LOW pH

Runs TAA053 through TAA064 were made to determine the effect of low, controlled scrubber inlet pH on SO_2 removal. All 12 runs were made with forced oxidation. The previous adipic acid runs with limestone stoichiometry controlled at 1.2 moles Ca/mole SO_2 absorbed, Runs TAA001 through TAA044, had demonstrated that pH had negligible effect on SO_2 removal at constant limestone stoichiometry and a high scrubber inlet pH of 5.6 to 6.1.

Runs TAA053 through TAA064 include data for four levels of scrubber inlet pH: 4.6, 5.0, 5.3, and 5.6. The pH = 5.6 level corresponds to a limestone stoichiometry of 1.2, and is thus a base case for comparison with the high-pH tests. The adipic acid concentrations tested are 600, 1300, and 2000 ppm. The slurry and gas flow rates are constant at 900 gpm and 25,000 acfm (at 300°F), respectively, and the total static height of spheres is 15 inches.

Equation 10-2 below has been fitted to the SO_2 removal data for Runs TAA053 through TAA064:

$$\begin{matrix}\text{Fraction } SO_2\\ \text{Removal}\end{matrix} = 1 - \exp[-4.7 \times 10^{-3} \exp(\text{pH} + 6.2 \times 10^{-4} A)] \qquad (10\text{-}2)$$

where

pH is that of the TCA inlet liquor, and A is the adipic acid concentration in ppm.

Equation 10-2 explains 98 percent of the variation in the SO_2 removal data for Runs TAA-053 through TAA-064, with a standard error of estimate of 2.2 percent SO_2 removal.

Figure 10-4 shows SO_2 removal as a function of adipic acid concentration and TCA inlet liquor pH. The lines in Figure 10-4 are predicted by Equation 10-2; the actual data points are also shown.

In comparing Figure 10-4 with Figures 10-2 and 10-3 (high pH data), the low pH curves of Figure 10-4 have a noticeably steeper slope for adipic acid concentration above 600 ppm than is the case for the high pH data. At low pH, the adipic acid is partially ineffective because of significant amounts of unionized adipic acid. For example, Figure 10-3 shows that, at 1.2 limestone stoichiometry (inlet pH of 5.6 to 6.1) and 28 gpm/ft^2 (900 gpm), 75 percent SO_2 removal can be achieved with zero adipic acid concentration. To achieve this same 75 percent SO_2 removal, Figure 10-4 shows that 600 ppm of adipic acid is required at 5.3 inlet pH, and 1700 ppm at 4.6 inlet pH.

Therefore, operation at low pH with limestone/adipic acid is undesirable in two respects: adipic acid is only partially utilized for scrubbing, and the SO_2 removal is far more sensitive to fluctuations in both pH and adipic acid concentration, resulting in poor system control.

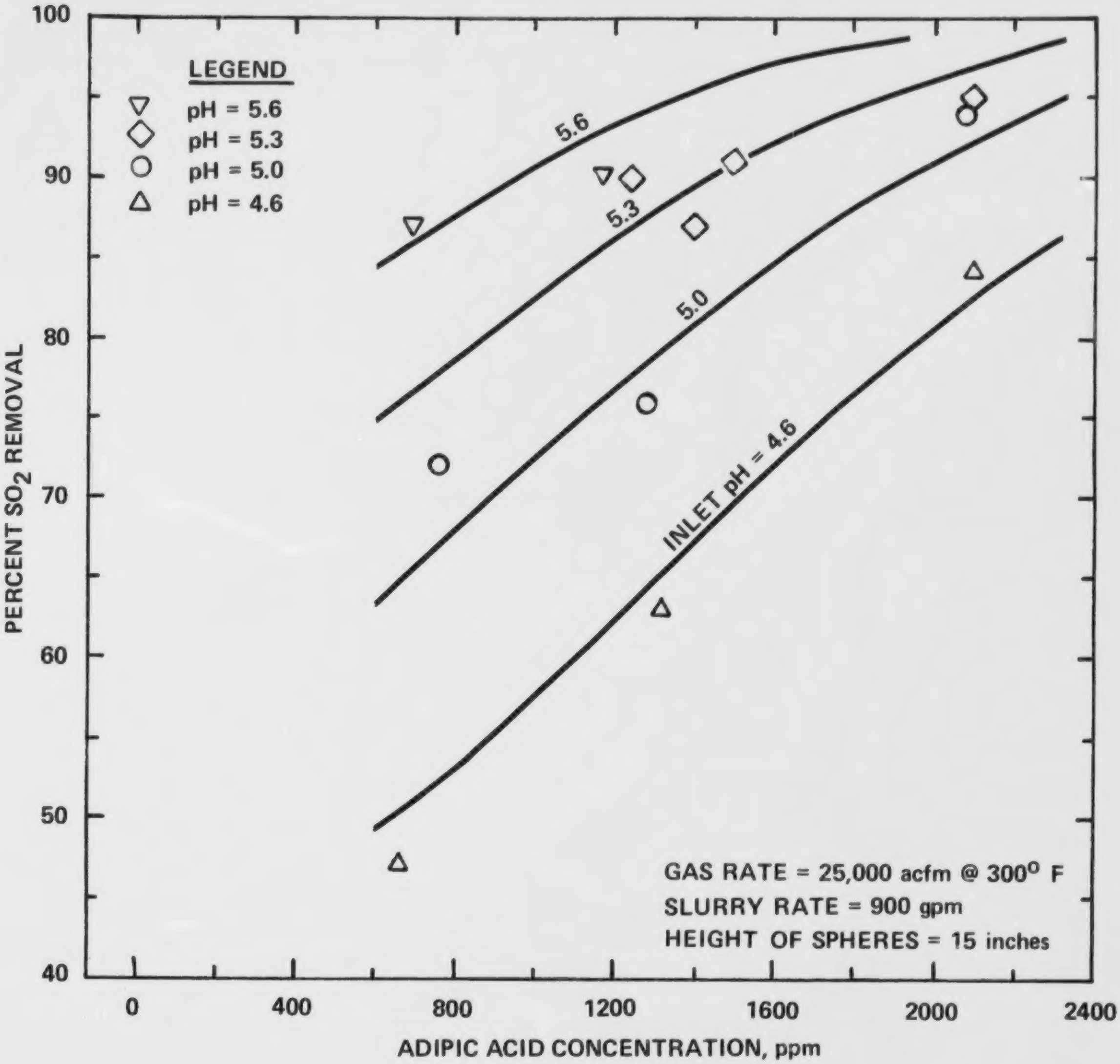


FIGURE 10-4

EFFECT OF ADIPIC ACID CONCENTRATION AND SCRUBBER INLET pH ON SO_2 REMOVAL IN THE TCA WITH FOUR GRIDS AND 15 INCHES OF SPHERES

Figure 10-5 is a plot of hourly SO_2 removal data as a function of scrubber inlet pH and adipic acid concentration for Runs TAA006, TAA025, TAA028, TAA032, and TAA053 through TAA064. The plot further demonstrates that the SO_2 removal is sensitive to both pH and adipic acid concentration for scrubber inlet pH below about 5.3-5.6.

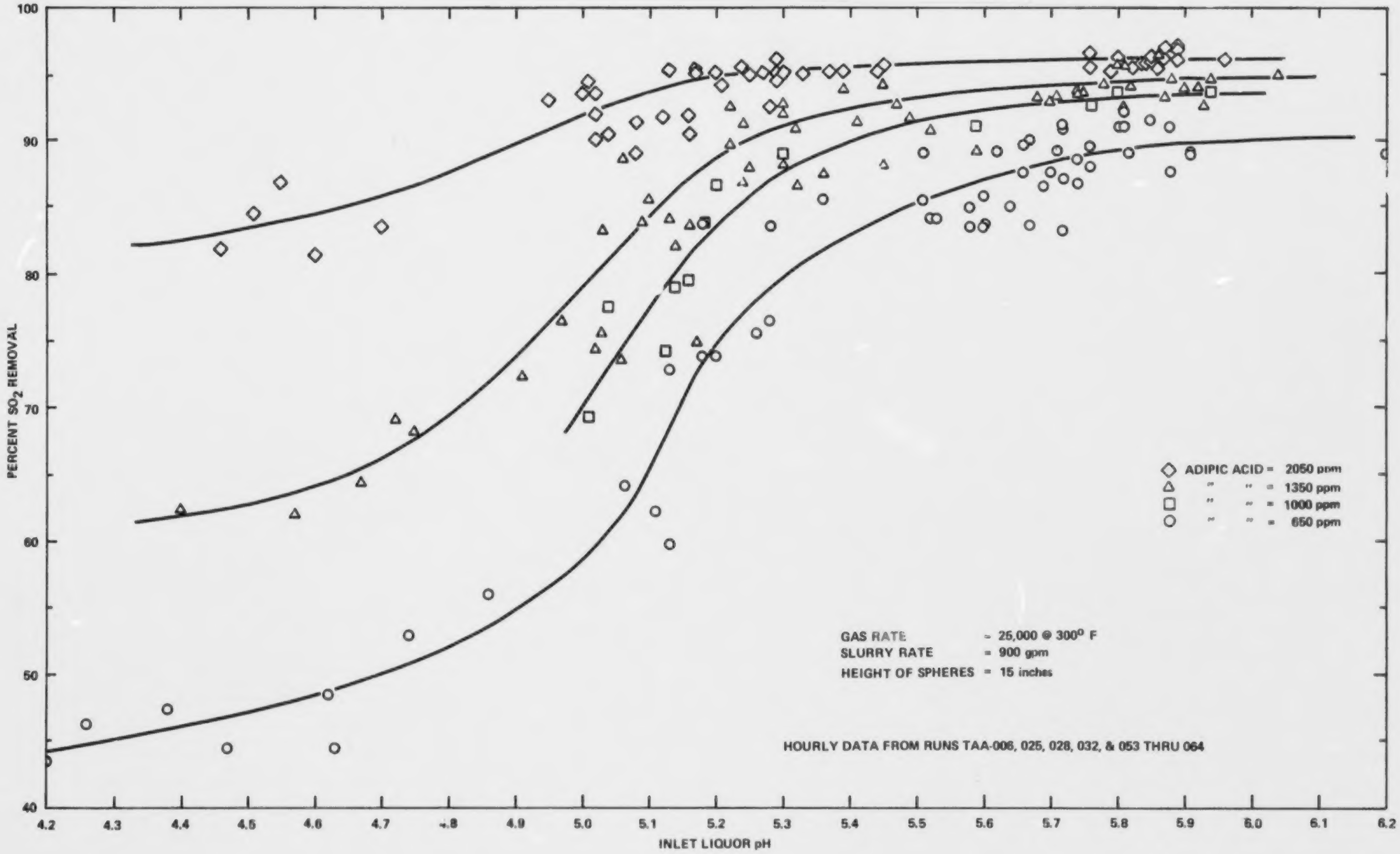


Figure 10-5. EFFECT OF SCRUBBER INLET pH AND ADIPIC ACID CONCENTRATION ON SO_2 REMOVAL

Section 11

OTHER TESTING ON THE TCA SYSTEM WITHOUT ADIPIC ACID ADDITIVE

The results of testing on the TCA system without adipic acid addition from November 1978 through October 1979 are briefly summarized in this section. The TCA system was turned over to TVA in October 1979 for modifications and preparation for a six-month demonstration of the DOWA basic aluminum oxide process.

LIME RUN 625-2A

A 1652-hour long-term reliability lime test with one tank and without forced oxidation was conducted on the TCA system from mid-November 1978 through late-January 1979.

As with the venturi/spray tower limestone/adipic acid long-term (1666 on-stream hours) reliability Run 907-1B (Reference 1), Run 625-2A was made for the EPA to assess continuous flue gas monitoring procedures. Another objective of the test was to continue the demonstration of the long-term operating reliability of the TCA lime scrubbing system.

The major test conditions and average test results are listed in Table 11-1. The flue gas flow rate was varied between 17,000 and 30,000 acfm (gas velocity between 7.1 and 12.5 ft/sec) based on a "typical" daily boiler load cycle. The slurry flow rate was constant at 1450 gpm (L/G = 60 to 106 gal/mcf). The scrubber inlet pH was controlled at 8. Three beds were used, with 5 inches static height of 1-7/8 inch diameter 11.5 gram nitrile foam spheres in each bed.

The scrubber onstream factor was 99.8 percent based on a total of 4 hours of unscheduled outages, excluding a 139-hour boiler outage and scheduled outages for routine inspection.

The average SO_2 removal for the run was 90 percent at 2350 ppm average inlet SO_2 concentration. The centrifuge was used in series with the clarifier. Average centrifuge cake solids content was 58 percent. Inspections after 388 hours and 1652 hours (end of run) showed that the mist eliminator was entirely clean and the scrubber was free of plugging and scaling. There was no increase in solids deposits on the scrubber internals during Run 625-2A.

Table 11-1

TCA LIME LONG-TERM RELIABILITY TEST

Major Test Conditions	625-2A
Fly ash loading	High
Gas rate, acfm @ 300°F	17,000-30,000(1)
Slurry flow rate, gpm	1450
Percent solids recirculated	8
TCA inlet pH controlled at	8.0
EHT (D-204) residence time, min.	3.4
EHT (D-204) level, ft	17
D-204 downcomer used	Internal
Lime addition point	DC/EHT(2)
Total static bed height, inches	15
Run-Average Results	
Start-of-Run date	11/10/78
Onstream hours	1652
Percent SO_2 removal	90
Inlet SO_2 concentration, ppm	2350
SO_2 make-per-pass, m-moles/l	4-12
TCA inlet liquor percent gypsum saturation	105
Percent sulfite oxidation	16
Percent lime utilization	87

Notes:

(1) Gas rate was varied in accordance with a "typical" daily boiler load cycle.

(2) On 11/27/78 lime addition point was moved from internal downcomer (DC) to EHT.

EFFECT OF OXIDATION TANK AGITATOR SPEED AND AIR STOICHIOMETRY ON SULFITE OXIDATION

The purpose of this test series was to investigate the effect of the oxidation tank agitator speed and the air stoichiometry on the degree of sulfite oxidation. Limestone slurry without adipic acid was used in the tests. A single tank was used to give a more sensitive oxidation response. The flow configuration and the arrangement of the oxidation tank are shown in Figures 7-1 and 7-2, respectively. Except for three runs in which the external downcomer for oxidation tank was used, all runs were made with the internal downcomer.

Table 11-2 lists the results of 30 runs made in this test series. Operating conditions common to all runs were:

Fly ash loading: High (3 to 6 grains/dry scf)
Gas flow rate: 30,000 acfm @ 300°F
Slurry flow rate: 1200 gpm
Slurry solids concentration: 15 percent
Limestone stoichiometry: 1.2 moles Ca/mole SO_2 absorbed (TCA inlet pH greater than 5.1)
Oxidation tank level: 17 ft
Total static bed height: 15 inches of 1-7/8 inch diameter nitrile foam spheres

Three levels of agitator speeds, 18, 36, and 54 rpm, were studied. Initially the air flow rate was fixed during a run (Runs 904-2A through 904-2G) to give the desired air stoichiometric ratio. However, because of the variation in inlet SO_2 concentration and the decreasing trend of the SO_2 concentration during the initial runs, the resulting air stoichiometry was often excessively high with the accompanying high sulfite oxidation. Therefore, beginning from Run 904-2H, the air flow rate was varied during the run to maintain a constant air stoichiometric ratio.

The effect of air stoichiometry and oxidation tank agitator speed on sulfite oxidation is plotted on Figure 11-1. Some scatter of data is observed. Four runs are not included in the plot: Runs 904-2F and 904-2G with air stoichiometries greater than 3, and Runs 904-2H and 904-2I with abnormally low inlet SO_2 concentrations (high expected natural oxidation level). In general, however, greater than 97 percent oxidation was achieved with air stoichiometry higher than about 1.25 at 54 rpm agitator speed, about 2.0 at 36 rpm, and about 2.5 at 18 rpm. Oxidation dropped off sharply with air stoichiometry lower than these levels.

High liquor sulfite accompanied by low limestone utilization (36 to 69 percent) was observed for 7 runs (Runs 904-2D, 904-2J, 904-2L, 904-2M, 904-2N, 904-2O, and 905-2E). These are indicated by the darkened symbols in Figure 11-1. Similar phenomenon (high sulfite supersaturation and limestone blinding) had been experienced during limestone/adipic acid tests with forced oxidation using one tank (Section 8).

F

Table 11-2

RESULTS OF TCA LIMESTONE TESTS WITH FORCED OXIDATION USING ONE TANK - EFFECT OF AIR STOICHIOMETRY AND AGITATOR SPEED

Major Test Conditions	904-2A	904-2B	904-2C	904-2D	904-2E	904-2F	904-2G	904-2H	904-2I
Fly ash loading	High	High	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000
Liquor rate, gpm	1200	1200	1200	1200	1200	1200	1200	1200	1200
Percent solids recirculated	15	15	15	15	15	15	15	15	15
Limestone stoich. ratio(1) (controlled)	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
Oxidation tank residence time, min.	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1
Tank level, ft.	17	17	17	17	17	17	17	17	17
Air rate to oxidizer, scfm	185	110	185	110	185	300	300	40(2)	40(2)
Oxidizer agitator speed, rpm	54	54	36	36	18	18	36	36	54
Run-Average Results									
Start-of-Run Date	6/26/79	6/3079	7/6/79	7/15/79	7/19/79	7/23/79	7/30/79	8/2/79	8/7/79
Onstream hours	97	124	144	94	100	46	71	125	40
Percent SO_2 removal	81.1	80.6	77.2	66.1	81.8	82.0	81.4	81.9	82.8
Inlet SO_2 concentration, ppm	2355	2165	2220	2315	1870	1965	1225	1165	1160
SO_2 make-per-pass, m-moles/l	10.1	9.2	9.2	8.3	8.1	8.3	5.3	5.1	5.2
Inlet liquor $SO_3^=$ concentration, ppm	244	226	96	1041	22	21	25	237	23
Inlet liquor percent gypsum saturation	115	110	110	155	110	95	95	110	90
Slurry percent solids	14.9	15.1	14.5	15.2	14.3	14.7	15.5	15.3	14.6
Percent sulfite oxidation	98.0	99.1	99.4	97.4	99.1	99.3	99.4	99.1	99.0
Percent limestone utilization	84	80	84	44	82	85	86	79	82
Inlet pH	5.50	5.40	5.30	5.15	5.25	5.40	5.55	5.40	5.30
Filter cake solids, wt %	85	86	88	81	86	86	-	-	-
Air stoichiometry, atoms O/ mole SO_2 abs.	2.01	1.29	2.24	1.48	2.49	3.81	6.10	0.85	0.85
Mist eliminator restriction, %	-	-	-	-	-	-	-	-	0

(1) pH underride of 5.2
(2) Run with air stoichiometry controlled at 1.0

Table 11-2 (Continued)

RESULTS OF TCA LIMESTONE TESTS WITH FORCED OXIDATION USING ONE TANK - EFFECT OF AIR STOICHIOMETRY AND AGITATOR SPEED

Major Test Conditions	904-2J	904-2K	904-2L	904-2M	904-2N	904-2O	904-2P	904-2Q	904-2R	904-2S	904-2T	904-2U
Fly ash loading	High	High	High	High	High	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000
Liquor rate, gpm	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200
Percent solids recirculated	15	15	15	15	15	15	15	15	15	15	15	15
Limestone stoich. ratio(1) (controlled)	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
Oxidation tank residence time, min.	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1(3)	4.1(3)	4.1(3)
Tank level, ft.	17	17	17	17	17	17	17	17	17	17	17	17
Air rate to oxidizer, scfm(2)	93	114	118	140	196	160	151	233	210	117	149	208
Oxidizer agitator speed, rpm	54	54	54	36	36	36	18	18	18	54	36	18
Run-Average Results												
Start-of-Run Date	9/4/79	9/6/79	9/7/79	9/8/79	9/8/79	9/10/79	9/11/79	9/12/79	9/13/79	9/15/79	9/16/79	9/17/79
Onstream hours	33	24	24	14	38	26	24	29	16	44	24	24
Percent SO_2 removal	79.9	84.6	78.7	75.8	78.6	73.9	74.8	75.7	81.6	83.7	81.6	85.6
Inlet SO_2 concentration, ppm	1858	1618	2065	2025	2226	2093	1998	2474	2292	1912	1767	2125
SO_2 make-per-pass, m-moles/l	7.9	7.3	8.6	8.1	9.3	8.2	7.9	9.9	9.9	8.5	7.7	9.7
Inlet liquor $SO_3^=$ concentration, ppm	722	7	174	669	366	470	158	249	285	122	100	226
Inlet liquor percent gypsum saturation	130	95	125	125	130	105	120	110	115	110	110	105
Slurry percent solids	15.0	14.9	14.8	15.9	14.1	14.8	13.2	15.0	13.8	15.7	14.6	9.0
Percent sulfite oxidation	98.0	99.7	98.5	99.2	98.0	98.1	76.2	74.7	75.1	99.0	98.4	69.5
Percent limestone utilization	44	81	44	36	60	60	68	62	66	73	75	21
Inlet pH	5.20	5.40	5.05	5.10	5.25	5.00	5.25	5.25	5.45	5.35	5.35	5.65
Filter cake solids, wt %	83	86	80	85	82	86	--	71	70	87	82	-
Air stoichiometry, atoms O/ mole SO_2 abs.	1.30	1.70	1.50	1.85	2.30	2.10	2.05	2.55	2.30	1.50	2.10	2.35
Mist eliminator restriction, %	-	-	-	-	-	-	-	-	-	-	-	-

(1) pH underride of 5.2
(2) Run with air stoichiometry control
(3) External downcomer for oxidation tank was used

Table 11-2 (Continued)

RESULTS OF TCA LIMESTONE TESTS WITH FORCED OXIDATION USING ONE TANK - EFFECT OF AIR STOICHIOMETRY AND AGITATOR SPEED

Major Test Conditions	905-2A	905-2B	905-2C	905-2D	905-2E	905-2F	905-2G	905-2H	905-2I
Fly ash loading	High	High	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000
Liquor rate, gpm	1200	1200	1200	1200	1200	1200	1200	1200	1200
Percent solids recirculated	15	15	15	15	15	15	15	15	15
Limestone stoich. ratio(1) (controlled)	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
Oxidation tank residence time, min.	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1
Tank level, ft.	17	17	17	17	17	17	17	17	17
Air rate to oxidizer, scfm(2)	0	53	67	87	116	117	128	148	172
Oxidizer Agitator Speed, rpm	54	54	54	54	54	36	36	36	36
Run-Average Results									
Start-of-Run Date	9/18/79	9/19/79	9/20/79	9/21/79	9/22/79	9/23/79	9/24/79	9/25/79	9/26/79
Onstream hours	20	20	24	24	24	23	24	24	24
Percent SO_2 removal	73.2	73.3	75.9	73.0	74.6	77.5	77.6	77.6	78.0
Inlet SO_2 concentration, ppm	2403	2545	2510	2280	2249	2226	2184	2238	2326
SO_2 make-per-pass, m-moles/l	9.3	9.9	10.1	8.8	8.9	9.2	9.0	9.2	9.6
Inlet liquor $SO_3^=$ concentration, ppm	36	231	222	72	602	190	240	150	172
Inlet liquor percent gypsum saturation	130	115	120	130	115	115	110	110	105
Slurry percent solids	15.3	15.8	16.1	14.3	15.0	14.9	16.0	15.4	15.1
Percent sulfite oxidation	35	55	59	62	85	67	69	66	65
Percent limestone utilization	83	80	78	80	69	83	86	83	81
Inlet pH	5.40	5.25	5.35	5.30	5.25	5.45	5.55	5.40	5.45
Filter cake solids, wt %	59	64	64	78	74	68	67	70	68
Air stoichiometry, atoms O/mole SO_2 abs.	0	0.60	0.72	1.06	1.45	1.40	1.55	1.75	1.95
Mist eliminator restriction, %	-	-	-	-	-	-	-	-	-

(1) pH underride of 5.2
(2) Run with air stoichiometry control

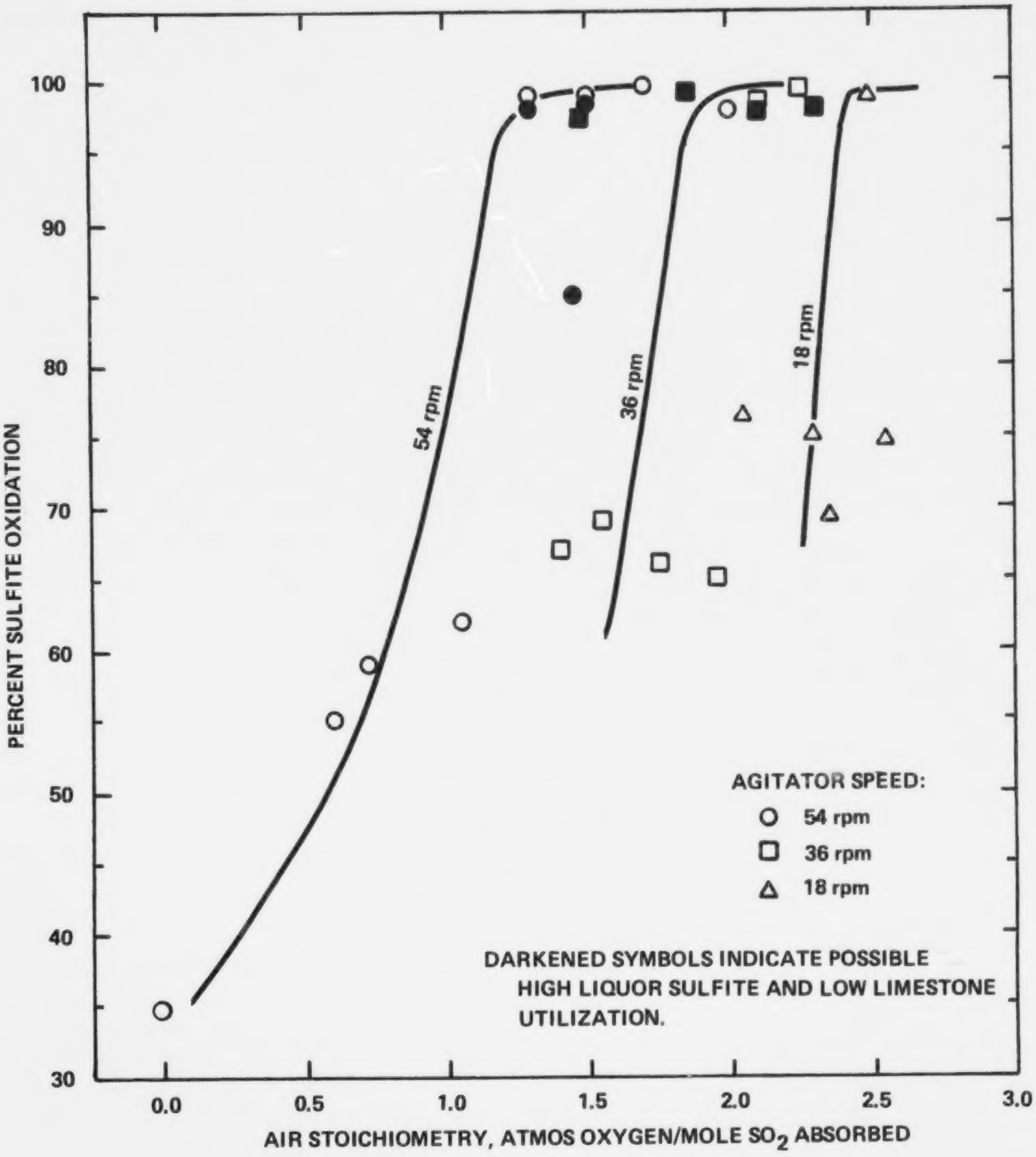


FIGURE 11-1
EFFECT OF AIR STOICHIOMETRY AND
AGITATOR SPEED ON SULFITE OXIDATION

Low limestone utilization was observed for most of the runs conducted at 18 rpm agitator speed (Runs 904-2P, 904-2Q, 904-2R, and 904-2U). Limestone utilization for these runs ranged from 21 to 68 percent with 70 to 76 percent oxidation. Poor agitation with the lower agitator speed probably resulted in slower limestone dissolution rate. Limestone blinding by $CaSO_3$ was not the cause because sulfite oxidation was low (sufficient $CaSO_3$ solid crystals for sulfite desupersaturation) and the liquor sulfite concentration ranged only 160 to 285 ppm.

LIMESTONE FACTORIAL TESTS WITH LOW INLET SO_2

A series of 14 limestone factorial tests without adipic acid were conducted during a period in which the inlet SO_2 concentration was too low (1360-2240 ppm) to obtain any meaningful data for comparison with other high sulfur results. Hopefully, the results obtained in this factorial series will serve as base cases for future testing when low inlet SO_2 concentration may be encountered again.

Table 11-3 lists the results of the 14 tests. All runs were made with forced oxidation using one tank (Figures 7-1 and 7-2). External downcomer of the oxidation tank was used. Common test conditions were:

Fly ash loading: High (3-6 grains/dry scf)
Flue gas flow rate: 25,000 acfm @ 300°F
Slurry solids concentration: 15 percent
Oxidation tank level: 17 ft
Air flow rate to oxidizer: 220 scfm
Agitator speed: 54 rpm

The SO_2 removal ranged from a low of 42 percent (Run TAA303 at 593 gpm slurry flow, 78 percent limestone utilization, and no spheres) to a high of 84.7 percent (Run TAA313 at 1197 gpm slurry flow, 64 percent limestone utilization, and 15 inches of spheres).

Table 11-3

RESULTS OF LIMESTONE FACTORIAL TESTS WITH FORCED OXIDATION USING ONE TANK -LOW INLET SO_2 CONCENTRATION-

Run No.	Gas Rate (acfm)	Liquor Rate (1) (gpm)	L/G gal/Mcf	Press. Drop (in. H_2O)	TCA Inlet Liquor pH		TCA Outlet Liq. pH	Ave $AD^=$ Conc (ppm)	Ave Cl^- Conc (ppm)	LS Utiliz.	Inlet SO_2 Concentration (ppm)		Percent SO_2 Removal		Air Rate (scfm)	Sulfite Oxid. (%)	Spheres (inches)	Replicate
					Range	Ave					Range	Ave	Range	Ave				
TAA301	25,000	894	45	2.3	5.42-5.68	5.58	5.4	0	1338	87	1360-1800	1590	46-52	48.1	220	98.8	0	A
TAA302	25,000	1183	59	2.7	5.44-5.68	5.54	5.6	0	3160	85	1680-1800	1769	59-63	61.9	220	98.1	0	
TAA303	25,000	593	30	1.9	5.66-5.70	5.68	5.5	0	2538	78	1840-2040	1941	38-43	42.0	220	98.8	0	
TAA304	25,000	875	44	2.2	5.56-5.58	5.57	5.3	0	2503	71	1920-2000	1954	54-55	54.6	220	99.3	0	A
TAA305	25,000	886	44	2.3	5.65-5.87	5.75	5.5	0	2459	76	1480-1720	1603	58-61	59.6	220	99.5	0	
TAA306	25,000	1179	59	2.8	5.58-5.77	5.68	5.6	0	2370	70	1800-2000	1918	68-71	69.6	220	99.5	0	
TAA307	25,000	597	30	1.9	5.79-5.86	5.83	5.4	0	3275	61	2200-2240	2225	44-45	44.8	220	99.1	0	
TAA308	25,000	907	45	6.5	5.11-5.30	5.22	5.1	0	2316	89	1740-2100	1976	58-69	61.2	220	99.6	15	B
TAA309	25,000	1203	60	7.5	5.25-5.38	5.31	5.0	0	3994	79	1720-1920	1856	80-84	82.2	220	99.5	15	
TAA310	25,000	593	30	5.5	5.29-5.46	5.39	4.9	0	3181	75	1880-2080	2033	56-59	57.1	220	99.4	15	
TAA311	25,000	893	45	6.4	5.20-5.36	5.30	5.1	0	3846	82	1960-2080	2038	70-73	71.3	220	99.4	15	B
TAA312	25,000	901	45	6.5	5.50-5.68	5.58	5.2	0	3373	63	1880-1960	1911	77-78	77.2	220	98.7	15	
TAA313	25,000	1197	60	7.6	5.25-5.54	5.41	5.2	0	4689	64	1840-2000	1914	83-86	84.7	220	99.6	15	
TAA314	25,000	586	29	5.3	5.13-5.52	5.29	5.0	0	4808	67	1880-2080	1958	59-61	59.5	220	99.4	15	

(1) External downcomer used.

Section 12

DEWATERING CHARACTERISTICS OF ADIPIC ACID-ENHANCED LIME/LIMESTONE SLURRY

Settling and dewatering characteristics of slurry solids are routinely monitored in the Shawnee Laboratory by cylinder settling tests and vacuum funnel filtration tests. A comparison of the results of these monitoring tests from July 1978 through October 1979 for lime and limestone slurry with and without adipic acid addition is presented in this section and summarized in Table 12-1. All of the test runs made in this period were with flue gas having high fly ash loading, corresponding to slurry solids comprised of approximately 40 percent fly ash (dry basis).

Cylinder settling tests are performed with slurries containing 8 percent solids (lime) or 15 percent solids (limestone) at room temperature in a 1000-ml cylinder containing a rake which rotates at 0.16 rpm. The initial settling rate and ultimate settled solids concentration are recorded as indices of dewatering characteristics. The initial settling rate is only a qualitative index of the solids settling properties. Design rates for sizing clarifiers must take into consideration the hindered settling rate as the solids concentrate. The ultimate settled solids from the cylinder test represent the highest solids concentration achievable in a settling pond.

Vacuum funnel filtration tests are performed in a Buchner funnel with a Whatman 2 filter paper under a vacuum of 25 in. Hg. The funnel tests correlate well with the Shawnee rotary drum vacuum filter when the filter cloth is not blinded but the funnel test cakes tend to have lower solids concentration.

Table 12-1 shows the effects of adipic acid on both oxidized and unoxidized lime and limestone slurries. The data reported are for a range of adipic acid concentration of 350 to 3500 ppm. All samples were with high fly ash loading in which about 40 percent of the slurry solids was fly ash.

As reported previously for limestone slurry (Reference 1), settling and filtration characteristics of oxidized slurry are superior to those of unoxidized slurry. The same trend exists with limestone slurry samples containing adipic acid. Table 12-1 does not show a similar trend for the lime slurry as for limestone slurry. Insufficient lime data and the lower initial solids concentration of 8 percent for the lime slurry may have masked this trend.

The average initial settling rate for oxidized limestone slurry decreased from 1.1 cm/min. without adipic acid to 0.6 cm/min. with adipic acid. However,

Table 12-1

COMPARISON OF SHAWNEE WASTE SLURRY DEWATERING CHARACTERISTICS WITH AND WITHOUT ADIPIC ACID ADDITION

Alkali	Fly Ash Loading(1)	Forced Oxidation	Adipic Acid (2)	Initial Solids Conc., wt %	Initial Settling Rate, cm/min.		Ultimate Settled Solids, wt %		Funnel Test Cake Solids, wt %	
					Avg.	Range	Avg.	Range	Avg.	Range
Limestone	High	Yes	Yes	15	0.6	0.3-0.9	70	59-83	69	59-77
Limestone	High	Yes	No	15	1.1	0.6-1.4	73	62-86	74	65-88
Limestone	High	No	Yes	15	0.2	0.1-0.3	52	42-69	55	49-73
Limestone	High	No	No	15	0.2	0.1-0.5	54	41-67	57	48-66
Lime	High	Yes	Yes	8	1.7	1.5-1.9	55	50-60	61	54-69
Lime	High	Yes	No	8	-	-	-	-	-	-
Lime	High	No	Yes	8	1.5	1.2-1.9	51	44-63	62	48-73
Lime	High	No	No	8	1.5	0.8-2.2	49	43-57	56	50-62

(1) Slurries with high fly ash loading contain about 40 percent fly ash in solids.
(2) Adipic acid concentration range is 350 to 3500 ppm.

this rate was still considerably higher than settling rates without forced oxidation. Without forced oxidation, the initial settling rate averaged 0.2 cm/min. with or without adipic acid.

Adipic acid had little or no effect on ultimate settled solids or funnel test cake solids. The limestone data indicated a slight decline in solids quality with adipic acid but the decline was small compared with the difference between limestone slurry with and without forced oxidation.

In addition to the vacuum funnel filtration tests, the filter cake solids from the rotary drum vacuum filter were monitored. During forced-oxidation operation with adipic acid-enhanced limestone, the rotary drum filter cake solids concentration ranged from 80 to 90 weight percent solids. This range is no different than that obtained when operating with oxidized sludge in the absence of adipic acid.

In summary, the only significant effect of adipic acid addition on dewatering characteristics of limestone slurry was a decline in initial settling rate of oxidized slurry. These observations agree generally with those at the IERL-RTP pilot plant. Although data are sparse for lime slurry, preliminary data indicated an insignificant effect of adipic acid.

Section 13

SUMMARY OF CHARACTERISTICS OF ADIPIC ACID AS A SCRUBBER ADDITIVE

Based on the results of testing to-date both at the IERL-RTP pilot plant and at the Shawnee Test Facility, the characteristics of adipic acid as a scrubber additive can be summarized as follows:

BENEFICIAL ASPECTS

- Adipic acid significantly enhances SO_2 removal. The optimum concentration range is 700 to 1500 ppm.
- For effective utilization of adipic acid, the minimum required scrubber inlet pH is only about 5.2. The corresponding limestone utilization at this low pH is normally about 80 percent or higher; thus the quantity of waste solids generated is reduced. High limestone utilization also contributes to reliable scrubber operation.
- With proper pH control and sufficient adipic acid concentration (sufficient buffer capacity), steady outlet SO_2 concentrations can be maintained even with wide fluctuations of inlet SO_2 concentrations.
- Adipic acid-enhanced limestone scrubbing has lower projected capital and operating costs than unenhanced limestone or limestone/MgO scrubbing (Reference 1).
- Adipic acid should promote use of less expensive and less energy-intensive limestone rather than lime.
- The effectiveness of adipic acid is not affected by forced oxidation and it can be used with both lime and limestone in systems with or without forced oxidation.
- The effectiveness of adipic acid is not adversely affected by chlorides as is the limestone/MgO process. Thus it is especially attractive for very tightly closed-liquor-loop operation.

- Adipic acid does not significantly affect the quality of unoxidized solids (settling rate/filterability) as can high magnesium ion concentrations.

- When used with lime, both good SO_2 removal and sulfite oxidation can be achieved in a single-loop scrubbing system using within-scrubber-loop forced oxidation.

NEGATIVE ASPECTS

- Adipic acid decomposition and adsorption on solids (Reference 13) requires adding up to 5 times that theoretically required (Reference 1). (However, consumption over the ranges anticipated has negligible economic impact.)

- Some decomposition products, such as valeric acid, have an unpleasant odor.

- The clarification (settling) rate for oxidized solids may be somewhat retarded.

OTHER CONSIDERATIONS

Toxicity. Other possible secondary environmental effects of adipic acid have not yet been fully determined. However, preliminary results from Level 1 Chemical and Bioassay analyses showed no measurable difference in toxicity of samples with and without adipic acid addition. These samples were taken in February 1979 from a limestone/adipic acid forced-oxidation run and a base case limestone run without forced oxidation. It should be noted that adipic acid is a food additive.

Limestone Blinding and Sulfite Scaling. Under certain operating conditions, high calcium sulfite supersaturation could occur in adipic acid-enhanced lime or limestone scrubbing systems. Although sulfite supersaturation is not restricted to adipic acid-enhanced systems, the presence of adipic acid increases this potential.

Several factors could contribute to high sulfite supersaturation in systems with or without adipic acid addition:

- Forced oxidation depletes solid $CaSO_3$ crystals in the slurry for sulfite desupersaturation.

- Insufficient oxidation intensity (low air stoichiometry and/or low agitator speed) in the liquid phase could cause sulfite supersaturation even though sulfite oxidation in the solids may be nearly complete.

- High SO_2 make-per-pass increases the sulfite supersaturation in the scrubber effluent.

- High pH.

Adipic acid buffers the pH drop across the scrubber, and therefore increases the potential of sulfite scaling at the bottom part of the scrubber, especially if the scrubber inlet pH is high. It also increases the SO_2 make-per-pass and similarly increases the sulfite scaling tendancy at the bottom of the scrubber. In the case of limestone scrubbing, blinding of limestone by calcium sulfite could occur, resulting in low pH and poor limestone utilization.

The phenomena of high sulfite supersaturation (high liquor sulfite concentration) and low limestone utilization (limestone blinding) have been observed both in the single-loop system with forced oxidation using one tank (Sections 7, 8, and 11) and in the second loop of a two-loop system with the recycle of oxidized slurry from the first to second loop (Section 5). In the former case, use of two tanks effectively eliminated the phenomena (Section 9).

Section 14

FUTURE EPA TEST PROGRAM FOR 1980

The EPA/IERL-RTP test program with adipic acid-enhanced scrubbing systems is continuing. The following activities are either proceeding or planned for 1980:

TESTING WITH ADIPIC ACID

- Additional limestone factorial tests at Shawnee on the venturi/spray tower system in a single-loop mode (one tank) to develop the interrelationship of SO_2 removal and operating parameters and to optimize the adipic acid concentration.
- Adipic acid testing at Shawnee with the spray tower only (venturi scrubber removed from the system) with limestone and with or without forced oxidation.
- Full-scale demonstration at a host utility site of adipic acid-enhanced limestone or lime scrubbing with options for forced oxidation or forced oxidation and solids dewatering study.
- Simulation of a full-scale operating plant at Shawnee on adipic acid-enhanced scrubbing.

TESTING WITHOUT ADIPIC ACID

- Additional limestone tests at Shawnee in a single-loop system with forced oxidation using one tank.
- Limestone type and grind tests at the IERL-RTP pilot plant and at Shawnee.
- Water management study at Shawnee.
- Testing at Shawnee to obtain design information on spray tower internals.

- Testing at Shawnee of alternative scrubber internal packings.

CONCURRENT PROGRAMS

- IERL-RTP pilot plant testing for Shawnee support.
- Studies by Aerospace Corporation of the handling and disposal characteristics of waste sludges produced at Shawnee with adipic acid-enhanced scrubbing.
- Evaluations by TVA of the economics of lime and limestone scrubbing with regard to forced oxidation, magnesium oxide addition, and adipic acid addition.
- Characterization at Shawnee of waste solids dewatering properties.
- Characterization of limestone type and grind by Martin-Marietta Laboratories.
- Solids clarification and thickening studies by Auburn University at Shawnee.
- Process pump seal water study at Shawnee.
- Instumentation control study at Shawnee.
- Flue gas emission characterization at Shawnee.
- Continued refinement of and addition to the TVA/Bechtel Design/Economic Computer Program.

Section 15

REFERENCES

1. Head, H.N., et al., "Recent Results from EPA's Lime/Limestone Scrubbing Programs -- Adipic Acid as a Scrubber Additive", presented at the Fifth Symposium on Flue Gas Desulfurization, Las Vegas, Nevada, March 5-8, 1979.

2. Borgwardt, R.H., "Significant EPA/IERL-RTP Pilot Plant Results", in Proceedings: Industry Briefing on EPA Lime/Limestone Wet Scrubbing Test Programs (August 1978), EPA-600/7-79-092, March 1979 (pp. 1-9).

3. Rochelle, G.T., "The Effect of Additives on Mass Transfer in $CaCO_3$ or CaO Slurry Scrubbing of SO_2 from Waste Gases", Ind. Eng. Chem. Fundam., Vol. 16, No. 1, pp. 67-75, 1977.

4. Rochelle, G.T., "Process Alternatives for Stack Gas Desulfurization by Throwaway Scrubbing", Proceedings of Second Pacific Chemical Engineering Congress, Vol. I, p. 264, August 1977.

5. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Summary of Testing through October 1974, EPA-650/2-75-047, (NTIS PB 244901), June 1975.

6. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Advanced Program, First Progress Report, EPA-600/2-75-050, (NTIS PB 245279), September 1975.

7. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Advanced Program, Second Progress Report, EPA-600/7-76-008, (NTIS PB 258783), September 1976.

8. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Advanced Program, Third Progress Report, EPA-600/7-77-105, (NTIS PB 274544), September 1977.

9. Bechtel National, Inc., EPA Alkali Scrubbing Test Facility: Advanced Program, Fourth Progress Report, to be published, Fall 1979.

10. Head, H.N., et al., "Results of Lime and Limestone Testing with Forced Oxidation at the EPA Alkali Scrubbing Test Facility", in Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, EPA-600/7-78-58a, (NTIS PB 282090), March 1978 (pp. 170-204).

11. Head, H.N., "Results of Lime and Limestone Testing with Forced Oxidation at the EPA Alkali Scrubbing Test Facility - Second Report", in Proceedings: Industry Briefing on EPA Lime/Limestone Wet Scrubbing Test Programs (August 1978), EPA-600/7-79-092, March 1979 (pp. 10-53).

12. Bechtel National, Inc., EPA Alkali Scrubbing Test Facility: Advanced Program, Final Report, to be published.

13. Radian Corporation, Further Study of Adipic Acid Degradation in FGD Scrubbers, draft final report for EPA Contract 68-02-2608, Task 72, October 5, 1979.

Appendix

CONVERTING UNITS OF MEASURE

Environmental Protection Agency policy is to express all measurements in Agency documents in metric units. In this report, however, to avoid undue costs or lack of clarity, English units are used throughout. Conversion factors from English to metric units are given below:

To Convert From	To	Multiply By
scfm (60°F)	nm3/hr (0°C)	1.61
cfm	m3/hr	1.70
°F	°C	(°F-32)/1.8
ft	m	0.305
ft/hr	m/hr	0.305
ft/sec	m/sec	0.305
ft^2	m^2	0.0929
ft^2/tons per day	m^2/metric tons per day	0.102
gal/mcf	l/m^3	0.134
gpm	l/min	3.79
gpm/ft^2	$l/min/m^2$	40.8
gr/scf	gm/m^3	2.29
in.	cm	2.54
in. H_2O	mm Hg	1.87
lb	gm	454
lb-moles	gm-moles	454
lb-moles/hr	gm-moles/min	7.56
lb-moles/hr ft^2	$gm\text{-}moles/min/m^2$	81.4
lb-moles/min	gm-moles/sec	7.56
psia	kilopascal	6.895

Blank pg 114

TVA/EPRI SHAWNEE COCURRENT SCRUBBER TEST RESULTS

L. J. Henson
Emission Control Development Projects
Office of Power
Tennessee Valley Authority
Muscle Shoals, Alabama

Presented for Presentation at
Fifth Industry Briefing Conference
Results of EPA Lime/Limestone Wet Scrubbing Test Programs
Sponsored by the U.S. Environmental Protection Agency
Royal Villa Motel in Raleigh, North Carolina 27711
December 5, 1979

Preceding page blank

TVA/EPRI SHAWNEE COCURRENT SCRUBBER TEST RESULTS

L. J. Henson
Emission Control Development Projects
Office of Power
Tennessee Valley Authority
Muscle Shoals, Alabama

ABSTRACT

The Tennessee Valley Authority (TVA) is actively involved in developing and demonstrating improved scrubber concepts and wet-scrubbing processes for removing sulfur dioxide (SO_2) from boiler flue gas. In 1976-1977 TVA and the Electric Power Research Institute (EPRI) jointly funded a 1-MW pilot test (Colbert pilot plant) of the cocurrent scrubber concept. Based upon promising results from the pilot test, EPRI funded TVA to design and operate a 10-MW prototype cocurrent scrubber at the Shawnee Test Facility to further evaluate and demonstrate the cocurrent scrubber concept. The primary advantages of this scrubber concept which were identified in the pilot tests were smaller physical size of the scrubber and the general arrangement of the scrubber equipment which have a major impact upon capital investment and operating cost.

The major objectives of the EPRI test program at Shawnee were to identify operating conditions for satisfactory removal of SO_2 and particulates from the flue gas using lime and limestone slurries, to demonstrate the reliability of the system, and to determine design parameters for scale-up to a commercial-scale unit. Following the EPRI test program, TVA has continued to operate the prototype scrubber to resolve problems with pluggage in the mist eliminator and reheater. Also, TVA has modified the scrubber to permit operation with forced oxidation.

During the EPRI test program, while operating at a scrubber superficial gas velocity of 27 ft/sec, an SO_2 removal efficiency of greater than 90% and a particulate removal efficiency of approximately 99% were achieved with lime or limestone absorbent. However, total system reliability was not demonstrated at this scrubber gas velocity because of solids deposition in the mist eliminator and reheater. Solids deposits in the mist eliminator and reheater have been eliminated during the TVA test program by operating the scrubber at a 20 ft/sec superficial gas velocity. The highlights of both EPRI and TVA prototype cocurrent scrubber test programs are presented in this paper.

TVA/EPRI SHAWNEE COCURRENT SCRUBBER TEST RESULTS

INTRODUCTION

Background

The Tennessee Valley Authority (TVA) conducted sodium carbonate, air-water, and lime/limestone-scrubbing tests with a cocurrent (cocurrent flow of flue gas and scrubbing slurry) pilot flue gas scrubber at the TVA Colbert Steam Plant in 1976-1977. These tests successfully demonstrated that the cocurrent scrubber is an effective sulfur dioxide (SO_2) scrubber and provided justification for additional testing of the cocurrent-scrubbing concept with a larger scrubber system. As a result, a 10-MW prototype cocurrent scrubber was installed in the idle Hydro-Filter scrubber train at the Shawnee Test Facility. The cocurrent scrubber test program at Shawnee was funded by the Electric Power Research Institute (EPRI) and had the following overall objectives:

1. Establish scale-up similarities and differences between the Colbert pilot scrubber and the Shawnee scrubber.
2. Demonstrate the operability and reliability of the Shawnee cocurrent scrubber.
3. Verify and quantify the advantages of a cocurrent scrubber.
4. Obtain data for scale-up to a full-scale facility.

Complete onsite support for the cocurrent test program at Shawnee was provided by TVA. This included all operating, maintenance, laboratory, and engineering personnel needed to support the test program. Radian Corporation, serving as a subcontractor, provided supplementary technical support to TVA. The EPRI cocurrent prototype scrubber test program at the Shawnee facility was completed in July. Based upon the promising results of the EPRI test program, TVA began to sponsor an additional test program with the cocurrent prototype scrubber.

The EPRI test program was composed of a series of test blocks which included air-water tests, sodium carbonate tests, lime-scrubbing variable studies, limestone-scrubbing variable studies, special air-slurry tests during a scheduled boiler outage, and long-term limestone-scrubbing reliability tests. The primary variables studied were tower height, gas velocity, slurry flow rate, slurry distribution, amount of tower packing, and absorbent stoichiometry. The limestone reliability tests in the TVA test program were continued to resolve the problem of solids deposition in the mist eliminator and reheater. Also, the scrubber system was modified for forced-oxidation testing.

Test Facility and Cocurrent Scrubber System Description

The Shawnee Test Facility is composed of three parallel 10-MW scrubbing systems: a venturi-spray tower system, Turbulent Contact Absorber (TCA) system, and the new cocurrent scrubber system. In addition to these individual scrubber trains, common facilities which support the three systems include limestone-grinding and slurry preparation equipment, a lime slaker for lime pebble, instrument air compressors, raw water booster pumps, and analytical laboratory and particulate laboratory facilities.

A schematic of the 10-MW cocurrent scrubber is shown in Figure 1 and the overall system flow diagram is shown in Figure 2. Tests are performed with the integrated system to simulate a closed-loop, full-scale facility. Flue gas is taken from either before or after the Shawnee Steam Plant No. 10 boiler precipitator and is drawn through the scrubber system by an induced draft (ID) fan. Normally the SO_2 concentration and particulate loading of the flue gas are 1,500-3,000 ppm and 3-5 gr/sft^3, respectively. An inline venturi meter ahead of the absorber continuously measures the gas flow.

Spray nozzles installed in headers at various levels in the scrubber distribute the lime or limestone slurry which is pumped from the scrubber circulation tank beneath the absorber. The nozzle arrangements for the spray headers are presented in Figure 3. There are also provisions for installing grids in the scrubber at 4- to 6-foot intervals beginning at about 3 feet below the top spray header. The scrubbed gas and the sprayed slurry flow downward through the absorber and enter the scrubber sump where they impinge on the surface of a pool of spent slurry. The gas turns 180 degrees upward and 90 degrees horizontally and passes through a mist eliminiminator and subsequently a steam tube reheater. Accumulated slurry drains from the scrubber sump to the scrubber circulation tank through a drain trap that maintains the sump liquid level and a vacuum seal on the scrubber system. Mist eliminator vanes are periodically washed with raw water or process liquor and the spent wash liquid drains to the scrubber circulation tank or the mist eliminator circulation tank. Steam from the No. 10 boiler superheater is used in the four soot blowers and the steam tube reheater. The reheater condensate returns by gravity to one of the boiler condensate return tanks. Flue gas leaving the reheater discharges through the ID fan and stack to the atmosphere. Concentrated slaked lime or limestone makeup slurry is continuously pumped to the scrubber circulation tank to control alkalinity in the system. Scrubber circulation slurry is pumped to the thickener to control a constant solids concentration in the scrubber circulation loop. The thickener overflow collects in a surge tank and is pumped back to the scrubber circulation tank or mist eliminator circulation tank. The thickener underflow is pumped to a horizontal vacuum belt filter for final dewatering of the sludge. The cake wash liquor and the filtrate from the filtrate receiver tanks are pumped to the scrubber circulation tank or the mist eliminator circulation tank. Makeup water is added to the system to replenish liquor lost in the filter cake and liquor required for flue gas saturation. The filter cake can be trucked to a disposal area or reslurried and pumped to a pond. Solids concentration in the scrubber circulation loop may be increased rapidly between test runs by transferring the thickener underflow directly to the scrubber circulation tank.

Cocurrent Scrubber Advantages

Advantages inherent to the cocurrent design which reduce capital investment and operating costs are:

- The equipment configuration is more compatible with most power plant duct and fan arrangements. The gas would enter the scrubber at a higher elevation and exit near ground level. The entrainment separator and reheat systems (likely to require the most maintenance) would be located near ground level. Likewise, the ID fans would be on the ground and the connecting ductwork to the stack would be shorter and probably less complex.
- The physical arrangement of the cocurrent scrubber causes the gas to change direction in the base of the unit before it enters the mist eliminator. Both the change in direction of the gas and the vertical position of the entrainment separator promote good liquid separation and drainage. Also, a separate mist eliminator wash loop can be used if needed.
- Scrubbing liquid should coalesce into larger droplets before disengaging from the gas stream near the base of the scrubber and further facilitate efficient operation of the mist eliminator.
- Flooding of the unit and associated high-pressure loss and excessive entrainment of scrubbing slurry (even if grids are added to improve gas-liquid contact) are less likely. Also, during normal cocurrent operation, the gas-side pressure loss would be lower since some liquid-side energy would be recovered.
- Higher gas velocities (smaller scrubbers) are achieved because of the reduced tendency to flood and because more efficient mist elimination is likely. Therefore, smaller or fewer scrubber modules would be required in a full-scale system.

SODIUM CARBONATE FACTORIAL TESTS

The EPRI cocurrent scrubber test program began in September 1978 with the air-slurry tests. During these tests, equipment checkout was completed and the equipment was operated as a totally integrated system. Also, special tests were performed to evaluate the reheater operation, measure the mist eliminator efficiency, and to determine the gas distribution in the scrubber.

Following the air-water tests, a series of tests using sodium carbonate as a reagent was performed during a 3-week period. The specific objectives of the sodium carbonate factorial tests were as follows:

- Evaluate system performance using a highly soluble alkaline additive to establish the maximum removal efficiency of the scrubber.
- Determine effects of major process variables on SO_2 removal and develop mathematical correlations.
- Provide information for development of lime-limestone factorial test plans.

Description of Tests

A total of 41 sodium carbonate factorial tests was conducted. The average length of each test was one shift at stabilized test conditions. Major operating conditions and ranges tested were as follows:

Variable	Range
Flue gas rate	12,000-25,000 aft^3/min at 300°F
Scrubber gas velocity	12.8-26.7 ft/sec at 125°F
Recirculated liquor rate	835-2,400 gal/min
Liquor-to-gas ratio, L/G	61-176 gal/kft^3
Scrubber height	25 and 45 feet
Distribution of liquor	Top of scrubber or equally through multiple spray headers

The factorial tests were divided into four blocks which corresponded to different arrangements of scrubber internals. The test blocks were organized according to spray header locations in service as follows:

Block 1 - Slurry distributed through spray header at 45 feet

Block 2 - Slurry distributed through spray header at 25 feet

Block 3 - Slurry distributed equally through spray headers at 15 and 25 feet

Block 4 - Slurry distributed equally through spray headers at 15, 25, 35, and 45 feet

The following operating conditions were held constant during the test block:

- Absorbent stoichiometry controlled at 1.0 mol Na_2CO_3/mol inlet SO_2 by adding 12% by weight sodium carbonate solution to the scrubber circulation tank.
- "Fly ash-free" flue gas.
- Flue gas quenched at first spray header.
- Makeup water added to control scrubber circulation tank level at 6 feet.
- Bleedstream from scrubber circulation tank to control sodium concentration in the recirculated liquor at 1% by weight.
- No grids in the scrubber.
- Scrubbed flue gas reheated to about 185°F.

A recirculated liquor pH of 8-9 was expected to provide a stoichiometry of 1.0 based upon Colbert pilot-plant test results. During Test Block 1, the recirculated liquor pH was held at 8. When it was observed that controlling the pH at 8 provided a sodium stoichiometry consistently greater than 1.0 mol Na_2CO_3/mol inlet SO_2, Test Blocks 2, 3, and 4 were operated with sodium stoichiometry control and the pH was monitored.

Test Results

Highlights of the major test conditions and results of the sodium carbonate factorial tests are presented in Tables I through IV. The overall SO_2 removal efficiency ranged from 84% to 99% as test conditions were varied. In the majority of the tests, the SO_2 removal efficiency was greater than 95%. Significant results from the sodium carbonate cocurrent scrubber tests were as follows:

- Increased scrubber residence time (provided by increasing spray header heights or decreasing gas velocities) provided higher SO_2 removals.
- An increase in liquor-to-gas ratio (L/G, gal/kft^3) resulted in an increase in SO_2 removal.
- Greater nozzle pressures increased SO_2 removal.
- Distribution of the liquor through multiple spray headers provided better SO_2 removal than distribution through a single header.

The observed effects of scrubber residence time (calculated by dividing the tower height by the gas velocity), L/G, and nozzle pressure on SO_2 removals are typical relationships which are expected for most scrubber systems and agree with the Colbert pilot-plant test results. The Colbert test results indicated an opposite relationship for the effect of distribution of liquor in the tower. This apparent discrepancy in spray distribution results for the 1-MW versus the 10-MW scrubbers may be explained by a difference in wall effects which was not accounted for during the tests.

Regression analysis of the Na_2CO_3 test results provided the following mathematical correlation of the major variables:

$$\ln (100 - \% \text{ removal}) = 4.279 - 0.874H/U - 0.053N - 0.000531L + 0.152D$$

H = tower height, feet

U = scrubber gas velocity, ft/sec

N = nozzle pressure, psi

L = liquor rate, gal/min

D = liquor distribution

Liquor distribution is designated as either +1 for all liquor distributed at the top of the scrubbing zone (Blocks 1 and 2) or -1 for liquor distributed equally through available spray headers (Blocks 3 and 4). H/U is scrubber residence time in seconds. The regression equation is applicable only for an open spray tower and a stoichiometry of 1.0 mol Na_2CO_3/mol inlet SO_2.

Conclusions

The conclusions resulting from the sodium carbonate tests are as follows:

1. SO_2 absorption in the scrubber is sufficient to achieve SO_2 removals greater than 95% with all of the spray header configurations tested.
2. Scrubber gas residence time, L/G, nozzle pressure, and liquor distribution had a significant effect upon SO_2 removal.
3. Observed SO_2 removals appeared to be constrained primarily by mass transfer area limitations. An important feature here may be losses of liquor to the scrubber walls.

4. Subsequent lime and limestone tests should be able to achieve adequate removals (≥90%) but performance at this level may be restricted due to mass transfer area limitations.

5. No special materials or equipment problems which would be expected to affect scrubber reliability were evident.

LIME TESTS

Lime Factorial Tests

The objectives of the lime factorial tests were to determine the effects of various operating conditions upon the SO_2 removal efficiency of the scrubber and to provide the basis for selection of operating conditions for the short-term lime reliability tests.

Description of Tests--

The original test program was designed to investigate the effects of liquor distribution in the scrubber, gas residence time in the scrubber, and slurry recirculation rate on SO_2 removal efficiencies at a constant stoichiometry of 1.0 mol Ca/mol inlet SO_2 and a superficial scrubber gas velocity of 27 ft/sec. Due to problems in controlling the scrubber system at a constant stoichiometry for all of the tests, stoichiometry became another variable for consideration. The average length of each test was 1 day. Major operating conditions and ranges for the test block were as follows:

Variable	Range
Flue gas rate	25,000 aft^3/min at 300°F
Scrubber gas velocity	26.7 ft/sec at 125°F
Recirculated slurry rate	800-2,400 gal/min
Liquor-to-gas ratio, L/G	40-120 gal/kft^3
Scrubber height	25, 35, and 45 feet
Distribution of slurry	Top of scrubber or equally through multiple spray headers
Stoichiometry	0.6-1.1 mols Ca/mol inlet SO_2
Grids	0-8
Depth of each grid	0-1-1/4 inches

The following test conditions were held constant during the test block:

- Flue gas with full loading of fly ash.
- Flue gas quenched at first spray header.

- Clarified liquor and excess makeup water mixed in recycle liquor surge tank and added to scrubber circulation tank to control level at 6 feet.
- Bleedstream from scrubber circulation tank controlled solids concentration at 10% by weight.
- Slaked lime slurry prepared with makeup water and added to scrubber circulation tank to control at desired stoichiometry.
- Sludge dewatered in thickener to approximately 30% solids by weight (horizontal-belt filter installation was not complete during these tests).
- Scrubbed flue gas reheated to about 185°F.

Test Results--

Highlights of the major test conditions and results of the lime factorial tests are presented in Tables V-A and V-B. The SO_2 removal efficiency ranged from 59% to 94% as test conditions were varied. The initial tests (L-400 and L-300) were performed without grids (open spray tower) and achieved less than 85% SO_2 removal. Consequently, essentially all of the remaining tests were performed with grids in the scrubber and with Bete full-cone slurry spray nozzles on a single header at the top of the scrubber. Five tests at the end of the test block (L-1400A through L-1500) were performed with open pipe nozzles instead of Bete spray nozzles. The percent SO_2 removal for these tests ranged from 69% to 84%.

Regression analysis of the test results provided the following mathematical correlation of the major variables:

$$\ln (100 - \%\ \text{removal}) = 6.67 - 2.38S - 0.587H/U - 0.00037L - 0.155D$$

S = lime stoichiometry, mols Ca/mol inlet SO_2

H = tower height, feet

U = scrubber gas velocity, ft/sec

L = recirculated slurry rate, gal/min

D = liquor distribution

Liquor distribution is designated as either +1 for all slurry distributed at the top of the scrubbing zone or -1 for slurry equally distributed through available spray headers. H/U is scrubber residence time in seconds. The regression equation is applicable only for a grid tower with two, 1-1/4-inch grids required for each 10-foot section of scrubber height and a recirculated slurry solids concentration of 10% by weight. The regression equation accounts for approximately 91% of the variation in the data. For this 91%, the stoichiometry accounts for 77%; the residence time, 6%; the liquor rate, 5%; and distribution, 3%. The remaining 9% is considered experimental error. For removal efficiencies greater than 80%, the maximum difference between the observed and predicted values is approximately 2%.

The regression equation was used to calculate the effect of the major variables. Examples of these calculations and comparisons with typical test results are as follows:

- An increase in stoichiometry from 1.0 to 1.2 mols Ca/mol inlet SO_2 for a 45-foot grid tower at a constant liquor rate (2,400 gal/min) and gas velocity (26.7 ft/sec) increased the SO_2 removal from 90.4% to 94.1%. Test L-401 achieved 94% SO_2 removal at a stoichiometry of 1.12 mols Ca/mol inlet SO_2 and a scrubber pressure drop of about 3.5 inches H_2O.

- An increase in stoichiometry from 1.0 to 1.2 mols Ca/mol inlet SO_2 for a 35-foot grid tower at a constant liquor rate (2,400 gal/min) and gas velocity (26.7 ft/sec) increased the SO_2 removal from 88.1% to 92.6%. Test L-500 achieved 91% SO_2 removal at a stoichiometry of 1.03 mols Ca/mol inlet SO_2 and a scrubber pressure drop of about 2.6 inches H_2O.

- An increase in grid tower height from 25 to 45 feet (increases gas residence time from 0.94 to 1.7 seconds) increases the SO_2 removal from 85.2% to 90.4% for a constant stoichiometry (1.0 mol Ca/mol inlet SO_2), liquor rate (2,400 gal/min), and gas velocity (26.7 ft/sec). Test L-600 achieved 85% SO_2 removal for a 25-foot grid tower and a scrubber pressure drop of about 1.9 inches H_2O.

- An increase in liquor rate from 800 to 2,400 gal/min (increases L/G from 40 to 120 gal/kft^3) in a 45-foot grid tower increases the SO_2 removal from 82.7% to 90.4% at a constant stoichiometry (1.0 mol Ca/mol inlet SO_2) and gas velocity (26.7 ft/sec). Test L-700 achieved 85% SO_2 removal for a liquor rate of 1,600 gal/min (L/G = 80 gal/kft^3) and a scrubber pressure drop of about 3.0 inches H_2O.

- An increase in liquor rate from 800 to 2,400 gal/min (increases L/G from 40 to 120 gal/kft^3) in a 35-foot grid tower increases the SO_2 removal from 78.5% to 88.1% at a constant stoichiometry (1.0 mol Ca/mol inlet SO_2) and gas velocity (26.7 ft/sec). Test L-800 achieved 82% SO_2 removal for a liquor rate of 1,600 gal/min (L/G = 80 gal/kft^3) and a scrubber pressure drop of about 2.3 inches H_2O.

Conclusions--

Conclusions based on the lime factorial tests were as follows:

- The SO_2 removal efficiency of the cocurrent open spray tower is less than 85% at the operating conditions tested in this program. Grids are required to provide adequate gas/liquor contact needed for 85-90% SO_2 removal.
- The variable that most strongly affects the SO_2 removal efficiency is the lime stoichiometry.
- The significance of the effect of other variables in descending order is gas residence time in the scrubber, liquor rate, and distribution of slurry in the scrubber (single versus multiple spray headers).
- Slurry distribution through a single spray header provides higher SO_2 removal than slurry distribution throughout the tower with multiple spray headers.
- The liquor rate required to achieve a desired level of SO_2 removal decreases as the scrubber gas residence time or grid tower height is increased at constant gas velocity and alkali stoichiometry. [For example, the regression equation predicts that for 27 ft/sec gas velocity and 1.0 mol Ca/mol inlet SO_2 alkali stoichiometry, the 25-foot, 35-foot, and 45-foot prototype grid towers require 2,370, 1,780, and 1,175 gal/min (L/G's of 119, 89, and 59 gal/kft^3), respectively, of recirculated liquor to achieve 85% SO_2 removal.]

Lime Short-Term Reliability Tests

The objective of the lime short-term reliability tests was to obtain more definitive information on the effect of scrubber operating conditions which provide at least 85% SO_2 removal upon the reliability of system components, such as the scrubber internals, mist eliminator, reheater, and ID fan.

Description of Tests--

Control of lime-scrubbing tests at a constant stoichiometry continued to be a problem. Recirculated slurry pH continued to vary from 6.0 to 11.0 for a stoichiometry of 1.0 mol Ca/mol inlet SO_2. In an attempt to stabilize pH, the excess alkalinity in the slurry was increased by increasing the recirculated slurry solids concentration from 10% to 15% solids by weight. The average length of each test was 4-5 days. Major operating conditions and ranges for the reliability tests were as follows:

Variable	Range
Flue gas rate	25,000 aft^3/min at 300°F
Scrubber gas velocity	26.7 ft/sec at 125°F
Recirculated slurry rate	1,200-2,400 gal/min
Liquor-to-gas ratio, L/G	60-120 gal/kft^3
Recirculated slurry solids concentration	10 and 15 weight percent
Scrubber height	35 and 45 feet
Distribution of slurry	Top of scrubber
Stoichiometry	1.0-1.1 mols Ca/mol inlet SO_2
Grids	6-8
Depth of each grid	1-1/4 inches

Other test conditions held constant were the same as for the lime factorial tests.

Test Results--

Highlights of the major test conditions and results of the lime short-term reliability tests are presented in Table VI. Essentially all of these tests had unstable operating conditions which make interpretation of the test results difficult. For tests L-2000, L-2001, and L-2002 with a 35-foot grid tower, a large percentage of the grid area was observed to be missing after the tests. The grids were probably damaged by large pieces of solids which fell from the solids deposits at the wet-dry interface. The displaced grids (possibly resulting in channeling of liquor in the tower) were a primary cause of the poor performance of the system during these tests. For the remaining tests, the tower height was increased to 45 feet so that the steam soot blower at the top of the scrubber could periodically clean the wet-dry interface. This arrangement prevented the buildup of large deposits of solids which were apparently falling onto and breaking the grids until improved grid supports could be installed.

During the last four short-term reliability tests the alkaline solids concentration in the recirculated slurry was increased by increasing the stoichiometry from 1.0 to 1.05 and 1.1 mols Ca/mol inlet SO_2 and the solids concentration in the recirculated slurry from 10% to 15% by weight. This increase of alkaline solids concentration in the slurry was made to increase SO_2 removal and to improve liquor pH control. (Alkalinity would be less sensitive to changes in inlet flue gas SO_2 concentrations and alkali feed rates.) During these last reliability tests, L-2200 through L-2500, significant improvements in percent SO_2 removals were observed for the higher stoichiometries (1.05 and 1.1 mols Ca/mol inlet SO_2) and recirculated slurry solids concentration (15% by weight) control points. The improvement in process control as a result of the higher alkaline solids concentration was not observed, however, until test L-2500.

Significant results from the lime short-term reliability tests were as follows:

- SO_2 removal was slightly lower than the removal achieved in Colbert pilot-plant tests. This difference was probably due to more slurry on the walls at Shawnee than at Colbert, tighter liquor loop closure at Colbert (55-60% solids) than at Shawnee (20-30% solids), and higher recirculated slurry solids concentration at Colbert (18%) than at Shawnee (15%).

- For a 45-foot grid tower, a scrubber pressure drop of about 3.0 inches H_2O, a stoichiometry of 1.1 mols Ca/mol inlet SO_2, an L/G of 60 gal/kft^3, and a recirculated slurry solids concentration of 15% by weight, the SO_2 removal efficiency is approximately 93% (test L-2400). For similar conditions in the Colbert tests approximately 96% SO_2 removal was achieved.

- While operating with a 45-foot grid tower, a scrubber pressure drop of about 3.5 inches H_2O, a stoichiometry of 1.05 mols Ca/mol inlet SO_2, an L/G of 90 gal/kft^3, and a recirculated slurry solids concentration of 15% by weight, the SO_2 removal efficiency ranged from 88% to 94%. The regression model, which is based upon a recirculated slurry solids concentration of 10% by weight, predicted an 89.5% SO_2 removal efficiency.

- For a 45-foot grid tower, the SO_2 removal efficiency was 91% for a scrubber pressure drop of about 3.5 inches H_2O, a stoichiometry of 1.0 mol Ca/mol inlet SO_2, and an L/G of 120 gal/kft^3 (test L-2100) which is the same as that predicted by the regression model.

- For a 45-foot grid tower, a scrubber pressure drop of about 3.5 inches H_2O, a stoichiometry of 1.1 mols Ca/mol inlet SO_2, an L/G of 120 gal/kft^3, and a recirculated slurry solids concentration of 15% by weight, the SO_2 removal efficiency ranged from 97% to 99% (test L-2300).

Conclusions--

The unstable operating conditions which were encountered during most of the lime short-term reliability tests (especially L-2000 through L-2400) prevented definite conclusions concerning the effects of operating conditions upon the scrubber efficiency and reliability. The tests, however, demonstrated trends in the scrubber removal efficiency and reliability as operating conditions were changed. Conclusions from these tests were as follows:

- The test results confirmed the levels of SO_2 removal efficiency of the cocurrent scrubber experienced during the lime factorial tests.

- A soot blower is needed to remove solids at the wet-dry interface.

- Pluggage in the mist eliminator and reheater for a superficial scrubber gas velocity of 26.7 ft/sec is a reliability problem.

- The SO_2 removal efficiency of the Shawnee prototype scrubber is slightly lower than the Colbert pilot scrubber. (While operating at L/G equal to 60 gal/kft^3, the SO_2 removal of the prototype unit was 93% for a recirculated solids concentration and stoichiometry of 15% by weight and 1.1 mols Ca/mol inlet SO_2, respectively. However, similar test conditions at Colbert achieved 96-97% SO_2 removal.) Possible causes are poorer liquor distribution or lower liquid-phase alkalinity in the Shawnee tests as compared to Colbert tests.

- The SO_2 removal efficiency of the prototype scrubber is increased slightly ($\sim$3%) when the alkalinity of the solids in the recirculated slurry is increased by increasing the solids concentration from 10% to 15% by weight.

- There was no significant change in the SO_2 removal efficiency of the scrubber or the tendency of the scrubber to form chemical scale when the scrubber system was operated on a closed liquor loop mode (20-30% solids by weight in the waste sludge) instead of open loop (bleed slurry sent to waste).

LIMESTONE TESTS

Limestone Factorial Tests

The purpose of the limestone factorial tests was to determine the effects of the various operating conditions upon the SO_2 removal efficiency of the scrubber and to provide the basis for selection of operating conditions for the short-term limestone reliability tests.

Description of Tests--

The limestone factorial test block was originally planned as a fractional factorial design which included 17 tests. However, preliminary evaluation of test results indicated that the 25-foot gas/liquor contact zone height was not sufficient for adequate SO_2 removal while operating at 27 ft/sec gas velocity. Consequently, the remaining tests with 25-foot scrubber height were eliminated and tests with 35-foot and 45-foot gas/liquor contact zone heights were added to the test block. The average length of each test was 1 day. The operating conditions which were varied and ranges tested were as follows:

Variable	Range
Flue gas rate	12,000, 20,000, and 25,000 aft^3/min at 300°F
Scrubber gas velocity	12.8, 21.3, and 26.7 ft/sec at 125°F
Recirculated slurry rate	1,000, 1,500, and 2,400 gal/min
Liquor-to-gas ratio, L/G	50-250 gal/kft^3
Scrubber height	25, 35, and 45 feet
Distribution of slurry	Top of scrubber or equally through multiple spray headers
Stoichiometry	1.0, 1.1, and 1.3 mols Ca/mol inlet SO_2
Grids	0-8
Depth of each grid	0-1-1/4 inches

The following test conditions were held constant:

- Flue gas with full loading of fly ash.
- Flue gas quenched at first spray header.
- Clarified liquor and excess makeup water mixed in recycle liquor surge tank and added to scrubber circulation tank to control level at 6 feet.
- Bleedstream from scrubber circulation tank controlled solids concentration at 15% by weight.
- Limestone slurry (60% solids by weight) prepared with makeup water and added to scrubber circulation tank to control at desired stoichiometry.
- Sludge dewatered in thickener to approximately 30% solids by weight.
- Scrubbed flue gas reheated to about 185°F.

Test Results

Highlights of the major conditions and results for the 25 limestone factorial tests are shown in Tables VII-A, VII-B, and VII-C. The tests are grouped according to physical configurations: scrubber height, slurry distribution (single or multiple headers), and open spray tower or grid tower.

Regression analysis of the test results provided the following mathematical correlation of the major variables:

$$\ln (100 - \%\ \text{removal}) = 11.0 - 0.978H/U - 0.00093L - 3.913S$$

S = limestone stoichiometry, mols Ca/mol inlet SO_2

H = tower height, feet

U = scrubber gas velocity, ft/sec

L = recirculated slurry rate, gal/min

H/U is scrubber residence time in seconds. The regression equation is applicable only for a grid tower with two, 1-1/4-inch grids required for each 10-foot section of scrubber height.

The regression equation accounts for 88% of the variation in the data. For this 88% the gas residence time accounts for 36%; the liquor rate, 29%; and the stoichiometry, 23%. The remaining 12% is considered experimental error.

Significant results based upon the limestone factorial test results and the regression equation were as follows:

- For liquor-to-gas ratios less than 95 gal/kft^3, the SO_2 removal efficiencies were 85% or less for all scrubber configurations and conditions tested.

- For an open spray tower with liquor-to-gas ratios of 120 and 150 gal/kft^3, SO_2 removal efficiencies were less than 83%. A grid tower is needed to provide gas/liquor contact required for adequate SO_2 removal efficiencies.

- A 25-foot gas/liquor contact zone height did not provide sufficient gas/liquor contact time required for adequate SO_2 removal efficiency (at 27 ft/sec gas velocity).

- The SO_2 removal efficiency of the scrubber increases as the limestone stoichiometry is increased to about 1.4 mols Ca/mol inlet SO_2. Higher stoichiometry does not significantly increase the removal efficiency.

Conclusions--

Conclusions based on the limestone factorial tests were as follows:

- A grid tower is needed for adequate SO_2 removal efficiencies.

- Three variables of those tested showed a significant effect on the SO_2 removal efficiency. These variables were residence time in the scrubber, liquor rate, and stoichiometry.

- Although the variables of nozzle pressure drop and distribution of the liquor throughout the tower were significant in the sodium carbonate test (performed with open spray tower), they did not significantly affect SO_2 removal in the limestone tests (with grid tower).

- For gas velocities greater than 25 ft/sec, the 25-foot grid tower height is not sufficient for SO_2 removal efficiency greater than 85%.

- A gas residence time of at least 1 second is required to obtain SO_2 removal efficiency greater than 85%.

- At the operating conditions tested, an alkali stoichiometry of approximately 1.0 mol Ca/mol inlet SO_2 is insufficient for an adequate SO_2 removal.

- For a 45-foot grid tower, high liquor rates (L/G's of approximately 90 to 100 gal/kft^3) are required to achieve 85-90% SO_2 removal at a gas velocity of about 27 ft/sec and with a stoichiometry of about 1.3 mols Ca/mol inlet SO_2.

- The high liquor-to-gas ratios needed to obtain adequate SO_2 removal indicate that the liquor distribution or gas/liquor contact in the grid tower is poor.

Limestone Short-Term Reliability Tests

The major objective of the limestone short-term reliability tests was to obtain more definitive information on the effect of scrubber operating conditions which provide at least 85% SO_2 removal upon the reliability of system components, such as scrubber internals, mist eliminator, reheater, and ID fan. The results of this test block were the basis for selecting the operating conditions for the limestone long-term reliability test.

Description of Tests--

The limestone factorials were performed with Fredonia limestone; however, the test facility switched to Stone Man limestone during the beginning of the short-term tests for other test program requirements. Therefore, the first test (LS-2100) was planned at the same conditions as factorial test LS-100 to determine the effect of the different limestone upon SO_2 removal efficiency. The facility switched back to Fredonia limestone at the end of test LS-2900. The scrubber train was inspected at the end of each test to determine the amount of solids deposition in the scrubber, mist eliminator, reheater, and ID fan. Major operating conditions and ranges for the reliability tests were as follows:

Variable	Range
Flue gas rate	17,000 and 25,000 aft^3/min at 300°F
Scrubber gas velocity	18.1 and 26.7 ft/sec at 125°F
Recirculated slurry rate	1,400-2,400 gal/min
Liquor-to-gas ratio, L/G	70-176 gal/kft^3
Recirculated slurry solids concentration	15 weight percent
Scrubber height	35 and 45 feet
Distribution of slurry	Top of scrubber
Stoichiometry	1.1 and 1.3 mols Ca/mol inlet SO_2
Grids	0-8
Depth of each grid	1-1/4 and 3-3/4 inches

Other test conditions held constant were the same as for the limestone factorial tests except for the use of the horizontal-belt filter to dewater the sludge to 60% to 85% solids by weight on March 20 during test LS-2400.

Test Results--

Highlights of the major test conditions and results are presented in Tables VIII-A and VIII-B. For all of the limestone short-term reliability tests, the inlet flue gas SO_2 concentration ranged from 1,800 to 3,000 ppm and the recirculated slurry pH normally ranged from 6.0 to 6.2. The first test, LS-2100, was with a 45-foot grid tower (1-1/4 inches/stage), a scrubber pressure drop of about 3.2 inches H_2O, a gas velocity of 26.7 ft/sec, a recirculated slurry rate of 2,400 gal/min (L/G = 120 gal/kft^3), and a stoichiometry of 1.3 mols Ca/mol inlet SO_2. The SO_2 removal efficiency of 93% to 95% for test LS-2100 confirmed the results of limestone factorial test LS-100 which had the same test conditions. This indicated that the use of Stone Man limestone instead of Fredonia limestone did not significantly change the SO_2 removal efficiency. For the next test, LS-2200, test conditions were the same as for the first test, LS-2100, except for replacement of the 90-degree slurry spray nozzles with 60-degree nozzles. The nozzle replacement was an attempt to improve the gas/liquor contact in the tower and reduce the high liquor-to-gas ratios needed to obtain adequate SO_2 removal. However, the SO_2 removal efficiency did not improve indicating no improvement in gas/liquor contact by decreasing the amount of slurry sprayed directly on the scrubber walls. The 60-degree nozzles were used for the remainder of the test program. For test LS-2300, the recirculated slurry rate was reduced from 2,400 gal/min (L/G = 120 gal/kft^3) to 2,000 gal/min (L/G = 100 gal/kft^3), and other test conditions remained constant. The SO_2 removal efficiency was 89-90% for a stoichiometry range of 1.21 to 1.44 mols Ca/mol inlet SO_2 and a scrubber pressure drop of about 3.0

inches H_2O. For test LS-2400, the grid tower height was reduced to 35 feet and the liquor rate increased to 2,400 gal/min (L/G = 120 gal/kft^3). During this test, the horizontal-belt filter was placed in operation. The SO_2 removal efficiency ranged from 88% to 91% for a range in stoichiometry of 1.25 to 1.40 mols Ca/mol inlet SO_2 and a scrubber pressure drop of about 3.0 inches H_2O.

Tests LS-2201 and LS-2900 were performed to permit Radian Corporation to make a study of gas-slurry distribution in an open tower and a grid tower. Test LS-2201 was at the same test conditions as for LS-2200 except for use of a horizontal-belt filter for dewatering the sludge. There was no detectable change in SO_2 removal efficiency (compared to test LS-2200) during test LS-2201 which operated for 38 hours. For test LS-2900, the grids were removed. For an inlet flue gas rate of 25,000 aft^3/min (scrubber gas velocity of 26.7 ft/sec), a scrubber pressure drop of 0.1 inch H_2O, and recirculated slurry rate of 2,400 gal/min (L/G = 120 gal/kft^3), the SO_2 removal efficiency ranged from 71% to 81% for a range in stoichiometry of 1.05 to 1.51 mols Ca/mol inlet SO_2. For test LS-2901, the inlet flue gas rate was decreased to 17,000 aft^3/min (gas velocity of 18.1 ft/sec) and the liquor-to-gas ratio increased to 176 gal/kft^3 (recirculated slurry rate of 2,400 gal/min). The SO_2 removal efficiency ranged from 87% to 92% for a range in stoichiometry of 1.07 to 1.37 mols Ca/mol inlet SO_2. The flue gas pressure increases 0.6 inch H_2O as it flows through the scrubber.

For tests LS-2700 and LS-2701 the depth of each of the six grids in the 35-foot grid tower was increased from 1-1/4 inches to 3-3/4 inches to determine the effect of thicker grids upon SO_2 removal. For a scrubber gas velocity of 26.7 ft/sec and a recirculated slurry rate of 1,800 gal/min (L/G = 90 gal/kft^3), the SO_2 removal efficiency ranged from 88% to 92% for a range in stoichiometry of 1.07 to 1.39 mols Ca/mol inlet SO_2. An SO_2 removal efficiency of 93% was achieved during test LS-2701 with a total grid depth of 22.5 inches (scrubber pressure drop of 7 inches H_2O), a stoichiometry of 1.3 mols Ca/mol inlet SO_2, and a recirculated slurry rate of 1,600 gal/min (L/G = 80 gal/kft^3). For test LS-2702, the recirculated slurry rate was reduced to 1,400 gal/min (L/G = 70 gal/kft^3) which provided a 90% SO_2 removal efficiency at a limestone stoichiometry of 1.3 mols Ca/mol inlet SO_2. Other test conditions were the same as for test LS-2701. Test conditions for test LS-2703 were the same as for test LS-2702 except for a change in the mist eliminator wash scheme in an attempt to improve mist eliminator and reheater reliability. For the final limestone short-term reliability test, LS-3000, the stoichiometry was reduced to 1.1 mols Ca/mol inlet SO_2 and the recirculated slurry rate was increased from 1,400 to 1,800 gal/min (L/G = 90 gal/kft^3) to obtain a 90% SO_2 removal efficiency. The flue gas pressure decreases about 7.2 inches as it flows through the scrubber. These test conditions were changed in an attempt to reduce the limestone in the entrained slurry entering the mist eliminator which could react with unremoved SO_2 and increase solids deposition in the mist eliminator and reheater. Also, the mist eliminator wash pressure was increased from 10 to 35 psi for this test. The operating time of this test was only 80 hours because it was interrupted by a boiler outage; consequently, the evaluation of the reliability of these operating conditions was not completed.

Key findings based upon the limestone factorial test results were as follows:

- A 90% SO_2 removal efficiency can be achieved with a 45-foot grid tower operating with a limestone stoichiometry of 1.3 mols Ca/mol inlet SO_2, a recirculated slurry rate of 2,000 gal/min (L/G = 100 gal/kft^3), a scrubber gas velocity of 27 ft/sec, and a scrubber pressure drop of about 3.0 inches H_2O. Eight, 1-1/4-inch-thick grids are required (test LS-2300).

- A 90% SO_2 removal efficiency can be achieved with a 35-foot grid tower operating with a limestone stoichiometry of 1.3 mols Ca/mol inlet SO_2, a recirculated slurry rate of 2,400 gal/min (L/G = 120 gal/kft^3), a scrubber gas velocity of 27 ft/sec, and a scrubber pressure drop of about 3.0 inches H_2O. Six, 1-1/4-inch-thick grids are required (test LS-2400).

- An 88% SO_2 removal efficiency can be achieved with a 45-foot open spray tower using a limestone slurry at a stoichiometry of 1.3 mols Ca/mol inlet SO_2, a scrubber recirculated slurry rate of 2,400 gal/min (L/G of 176 gal/kft^3), and a scrubber gas velocity of 18 ft/sec (2.5-second gas residence time). At this operating condition, the flue gas pressure increases 0.6 inch H_2O as it flows through the scrubber (test LS-2901).

- A 90% SO_2 removal efficiency can be achieved with a 35-foot grid tower operating with a limestone stoichiometry of 1.3 mols Ca/mol inlet SO_2, a recirculated slurry rate of 1,400 gal/min (L/G of 70 gal/kft^3), a scrubber gas velocity of 27 ft/sec, and a scrubber pressure drop of about 6.5 inches H_2O. Six, 3-3/4-inch-thick grids are required (tests LS-2702 and LS-2703).

- An 88% SO_2 removal efficiency can be achieved with a 35-foot grid tower operating with a limestone stoichiometry of 1.1 mols Ca/mol inlet SO_2, a recirculated slurry rate of 1,800 gal/min (L/G of 90 gal/kft^3), a scrubber gas velocity of 27 ft/sec, and a scrubber pressure drop of about 7.2 inches H_2O. Six, 3-3/4-inch-thick grids are required (test LS-3000).

Conclusions--

Conclusions based upon the limestone short-term reliability tests in the EPRI cocurrent test program were as follows:

- The primary reliability problem area in the cocurrent prototype scrubber is pluggage in the mist eliminator and reheater. None of the scrubber operating conditions or mist eliminator wash schemes tested have provided reliable operating conditions.

- The use of Stone Man limestone instead of Fredonia limestone did not significantly change the SO_2 removal efficiency.

- Distribution of scrubber recirculated slurry through 60-degree spray nozzles instead of 90-degree spray nozzles did not significantly change SO_2 removal efficiency. Apparently there was no significant improvement in the gas/liquor contact in the prototype scrubber when this attempt to decrease the amount of slurry sprayed directly on the scrubber walls was made.

- There was no detectable change in SO_2 removal efficiency by the scrubber when the horizontal vacuum belt filter was placed in service and the liquor loop closure was tightened.

Special Tests During Power Plant Boiler Outage

A series of tests to determine the distribution of gas and slurry in the scrubber, to evaluate mist eliminator efficiency, and to refine the horizontal-belt filter operation was performed during the scheduled unit 10 boiler outage which began on May 4 and ended on July 4.

The nine air-slurry tests were performed with different internal scrubber arrangements used in the lime-limestone test blocks and for revised internal scrubber designs in attempts to improve slurry distribution in the tower. Mist eliminator efficiency testing was performed simultaneously with the air-slurry tests to develop a direct entrainment measurement procedure which may be used to evaluate mist eliminator efficiency. Another major objective of these tests was to determine design changes which would eliminate pluggage in the mist eliminator and reheater. The scrubber sump partition was extended about 3 feet; a 45-degree baffle was installed in the 90-degree duct bend upstream of the mist eliminator housing; and mist eliminator sump baffles were installed. Throughout the air-slurry tests, mist eliminator and reheater pluggage was insignificant. Seventeen filtration tests with unoxidized slurry and one test with oxidized slurry from the TCA scrubber system were performed to refine operating conditions of the horizontal-belt filter. Laboratory filtration tests were also performed in conjunction with the horizontal-belt filter tests.

Major conclusions from the tests performed during the power plant boiler outage are as follows:

- The inlet gas was not equally distributed at the top of the tower; however, no distinctive maldistribution pattern was observed.

- The gas distribution leaving the scrubber sump and entering the mist eliminators is very poor. A swirling gas flow pattern was found in the vertical duct upstream of the 90-degree bend at the inlet of the mist eliminator housing (see Figure 4).

- The gas distribution entering the mist eliminators is greatly distorted. Localized gas velocities entering the mist eliminator are as high as 50-60 ft/sec which exceed the maximum recommended velocity of 30-40 ft/sec for acceptable performance of the mist eliminator (see Figure 4).

- The horizontal-belt filter achieved filtration rates up to 1.9 pounds of solids/ft^2 of filter area/minute of total cycle time based upon a 1/2-inch-thick filter cake with 60-80% solids by weight, a vacuum of 18-26 inches Hg, and a maximum feed rate of 30 gal/min of slurry with 23-30% solids by weight. The maximum feed rate was limited to 30 gal/min by the capacity of the auxiliary equipment.

- A horizontal-belt filter cake concentration of 90% solids by weight can be achieved with oxidized limestone scrubber slurry.

Limestone Long-Term Reliability Tests

Following the scheduled boiler outage, the limestone long-term reliability demonstration test began on July 5 and was completed on July 23. The objective of this test was to determine if the cocurrent scrubber can meet emission standards with no scaling and/or plugging in the scrubber system. The scrubber train was inspected during and at the end of the test to determine the amount of solids deposition in the scrubber, mist eliminator, reheater, and ID fan. After the reliability test, the final test, LS-4001, in the EPRI cocurrent scrubber test program was performed at a lower limestone stoichiometry (1.10 versus 1.30 mols Ca/mol inlet SO_2 used in the reliability test) to determine if lower stoichiometry decreased solids deposits in the scrubber train.

Description of Tests--

Test conditions for the limestone long-term reliability test, LS-4000, were as follows:

- A flow rate of 25,000 aft^3/min (300°F) of flue gas with a full loading of fly ash at a scrubber gas velocity of 26.7 ft/sec (125°F).

- Final scrubber recirculated slurry rate of 1,800 gal/min (L/G = 90 gal/kft^3) quenched flue gas at first spray header.

- Thirty-eight-foot grid tower with six scrubber grids which are 3-3/4 inches thick.

- The extended divider in the scrubber sump, the 45-degree baffle in the 90-degree bend upstream of the mist eliminator housing, and the mist eliminator sump baffles were not removed following the air-slurry tests.

- Clarified liquor and excess makeup water mixed in recycle liquor surge tank and added to scrubber circulation tank to control level at 6 feet.

- Bleedstream from scrubber circulation tank controlled solids concentration at 15% by weight.

- Limestone slurry (60% solids by weight) prepared with makeup water and added to scrubber circulation tank to control at stoichiometry of 1.3 mols Ca/mol inlet SO_2.

- Sludge dewatered in thickener and horizontal-belt filter to a solids concentration approximately 60% by weight.

- A first-stage, three-pass Heil chevron mist eliminator followed by a second-stage, four-pass Heil chevron mist eliminator removed entrained moisture.

- Scrubbed flue gas reheated to about 185°F.

For this test, a fifth nozzle was installed in the center of the four existing mist eliminator spray nozzles at all three wash locations. The mist eliminator wash rate was then increased from 24 to 30 gal/min to maintain a nozzle pressure of 35 psig. (The wash scheme is presented in the Mist Eliminator Performance section.)

After the limestone stoichiometry was reduced from 1.3 to 1.1 mols Ca/mol inlet SO_2 for the final test, LS-4001, the liquor-to-gas ratio was adjusted from 90 to 120 gal/kft^3 in order to assure a 24-hour average of 90% SO_2 removal.

Test Results--

Highlights of the major conditions and results for tests LS-4000 and LS-4001 are presented in Table IX. After the recirculated slurry rate was increased from 1,600 to 1,800 gal/min for test LS-4000, the SO_2 removal efficiency ranged from 85% to 94%. Daily averages of the SO_2 removal efficiency were greater than 90%. The limestone long-term reliability test, LS-4000, was terminated after 349 hours of operation because of excessive pressure drops in the scrubber, mist eliminator, and reheater.

The final test, LS-4001, was discontinued on July 31, the last day of the EPRI test program, after 182 hours of operation due to excessive pressure drop across the grid tower (increased from 8.0 to 12.0 inches H_2O). Upon inspection, the cross-sectional area at the wet-dry interface was found to be restricted about 60%. This problem occurred frequently during the short-term tests for a 35-foot tower. For a 45-foot tower, a steam soot blower is positioned to clean the wet-dry interface. These solids deposits should not be a problem in full-scale operation if a soot blower is located appropriately to clean the wet-dry interface. The SO_2 removal efficiency ranged from 90% to 92% for a range in stoichiometry of 1.06 to 1.09 mols Ca/mol inlet SO_2 and a recirculated slurry rate of 2,000 gal/min (L/G = 100 gal/kft^3). For a recirculated rate of 2,400 gal/min (L/G = 120 gal/kft^3), the SO_2 removal efficiency increased to a range from 94% to 96% with a range in stoichiometry of 1.03 to 1.12 mols Ca/mol inlet SO_2.

Conclusions

Conclusions from tests LS-4000 and LS-4001 are as follows:

- The prototype cocurrent scrubber operation continued to be unreliable primarily because of solids deposition in the mist eliminator and reheater. The solids deposits are suspected to be caused by poor distribution of gas leaving the scrubber and entering the mist eliminator. The poor gas distribution increases the amount of entrained slurry leaving the scrubber, unevenly distributes entrained slurry into the mist eliminator, distorts the mist eliminator wash spray pattern, and causes high localized velocities in the mist eliminator which may reentrain the slurry into the reheater.

- Soot blowing at the wet-dry interface is necessary to prevent a buildup of solids deposits. A presaturator should be installed at the inlet of the prototype tower where a soot blower is located.

- Operation with a lower limestone stoichiometry (1.10 versus 1.30) does not prevent pluggage in the mist eliminator and reheater. After the gas distribution in this area is improved, the effect of limestone stoichiometry upon the degree of solids deposition should be reinvestigated.

- The cocurrent prototype scrubber can consistently average 90% SO_2 removal efficiency over a 24-hour period at an inlet SO_2 concentration of approximately 2,500 ppm.

TVA Mist Eliminator Reliability Tests

The EPRI test program was concluded with the solids deposition problem in the cocurrent mist eliminator and reheater unresolved. Consequently, TVA decided to sponsor about 5 months of additional testing in an attempt to solve these problems and to perform tests with forced oxidation in the cocurrent scrubber system.

Description of Tests--

Major operating conditions and ranges tested were as follows:

Variable	Range
Flue gas rate	18,750 and 25,000 aft^3/min at 300°F
Scrubber gas velocity	20 and 26.7 ft/sec at 125°F
Recirculated slurry rate	1,400-1,900 gal/min
Liquor-to-gas ratio, L/G	93-100 gal/kft^3
Recirculated slurry solids concentration	10 weight percent
Scrubber height	38 feet
Distribution of slurry	Top of scrubber and presaturator at A-header location
Stoichiometry	1.3 mols Ca/mol inlet SO_2
Grids	6
Depth of each grid	3-3/4 inches

The major changes in operating conditions as compared to the EPRI reliability tests were lower scrubber superficial gas velocity, low-fly ash loading in the inlet flue gas [taken downstream of the unit 10 electrostatic precipitator (ESP)], and lower solids concentration in the scrubber recirculated slurry (10% versus 15%). Also, the scrubber height was increased to 38 feet due to the extension of the sump divider during the air-slurry tests. For some of the TVA tests, a presaturator was installed. Other conditions held constant are the same as for the EPRI limestone reliability tests.

Test Results

Highlights of the major test conditions and results are presented in Table X. The first TVA limestone mist eliminator reliability test, LS-5000C, began on August 8 and was designed to determine if reduction of the scrubber superficial gas velocity from 27 ft/sec to 20 ft/sec and operation with low-fly ash loading in the flue gas would eliminate solids deposition in the mist eliminator and reheater. The upstream face of the first-stage, Heil, three-pass chevron mist eliminator separator was washed with 30 gal/min of recycle liquor (1.7 gal/min/ft^2) for 1 minute and 40 seconds followed by 30 gal/min of raw water for 20 seconds. There was a delay of approximately 3 minutes between the raw water wash and recycle liquor wash. The downstream side of the first-stage and the upstream face of the second-stage Heil, four-pass chevron mist eliminator were washed with 30 gal/min of raw water for 30 seconds every 30 minutes. Five wash spray nozzles (1/2HH35WSQ Spraying Systems) were used for all three wash locations; positioned 7 to 11 inches from the mist eliminator; and operated at 35 psig. All of the spent wash liquor was drained to the scrubber circulation tank. After 319 hours of operation, solids depositions in the mist eliminator and reheater were insignificant; however, solids deposits at the wet-dry interface and in the scrubber outlet duct (upstream of mist eliminator) were substantial.

The next test, LS-5001C, at 20 ft/sec superficial gas velocity in the scrubber was designed to eliminate the wet-dry interface solids deposits observed in test LS-5000C and to further demonstrate the reliability of the entrainment separator and reheater. For this test two, 1-inch Bete spray nozzles were installed at the top of the scrubber tower (A-header) immediately below the steam soot blower. Approximately 95 gal/min of scrubber slurry was sprayed into the hot flue gas (280-310°F) to presaturate and cool it before it flowed into the scrubbing zone. This increased the total scrubber recirculated slurry from 1,400 gal/min to 1,495 gal/min and the liquor-to-gas ratio from 93 gal/kft^3 to 100 gal/kft^3. The scrubber walls at the wet-dry interface and the presaturator spray nozzles were cleaned once each shift with the soot blower which was operated with 150 psig steam. The other major operating conditions were the same as for test LS-5000C.

After 470 hours of operation, the system was inspected. The flow area restriction at the wet-dry interface was reduced from 50% to 13%. Flow area restriction at B-header was about 12%. The mist eliminator and reheater were essentially clean. Following test LS-5000C, the first-stage mist eliminator was cleaned but the second-stage mist eliminator and reheater were not. Therefore, after 789 hours of total operation the second-stage mist eliminator and reheater were proving reliable. For test LS-5001C, the SO_2 removal efficiency was consistently greater than 92%.

The next test, LS-5002C, at 20 ft/sec was designed to evaluate a new mist eliminator wash scheme based upon makeup water available for washing if the test was operated under a forced-oxidation test mode. The other major operating test conditions were the same as for test LS-5001C. After 348 hours of operation, the wet-dry interface, which was not cleaned following the previous tests, had not changed. Solids deposits at B-header had bridged the gap between opposing pairs of nozzles resulting in a 20% flow area restriction. The mist eliminator was 3-5% restricted and the reheater was clean. The solids deposits at the wet-dry interface are believed to be caused by inadequate gas/liquor contact with only two presaturator nozzles.

For the next test, LS-5010C, the scrubber gas velocity was increased to 26.7 ft/sec to compare operation with low-fly ash loading to previous operation with a full loading of fly ash in the flue gas. Major operating conditions were the same as for test LS-4000. The system was inspected after 238 hours of operation. The wet-dry interface flow area was 10% restricted and a 3% flow area restriction was observed at B-header location. The first-stage mist eliminator and reheater were plugged with solids and the second-stage mist eliminator was essentially clean.

Gas distribution measured during test LS-5000C (20 ft/sec in tower) was similar to that obtained during the EPRI test program (27 ft/sec in tower) except the maximum localized gas velocities had decreased from 50-60 ft/sec to 35-40 ft/sec. When the scrubber gas velocity is reduced from 27 to 20 ft/sec for the 38-foot grid tower with six scrubber grids which are 3-3/4 inches thick, the scrubber pressure drop decreases from about 7.2 to 3.5 inches H_2O.

Conclusions--

Conclusions based upon the TVA mist eliminator reliability tests are as follows:

- Reduction of the scrubber superficial gas velocity from 27 ft/sec to 20 ft/sec greatly reduces solids deposition in the mist eliminator and reheater. Apparently, operation at the lower velocity reduces the amount of entrainment leaving the scrubber and improves the mist eliminator efficiency.
- The overall general gas distribution entering the mist eliminator is similar for operation at scrubber velocities of 20 and 27 ft/sec. However, maximum localized gas velocities through the mist eliminator are reduced from 50-60 ft/sec to 35-40 ft/sec. The vendor claims that the mist eliminator can operate with gas velocities up to 30-40 ft/sec without reducing the efficiency of the mist eliminator.
- The reduction of solids deposition in the mist eliminator and reheater for a lower scrubber gas velocity is not primarily a result of operation with "fly ash-free" flue gas.
- A soot blower is necessary to remove solids at the wet-dry interface even for a presaturator with a low slurry rate.

Equipment Performance

Scrubber Performance--

Reliability of the cocurrent scrubber tower was not a problem during the test program except for solids deposition at the wet-dry interface. When slurry was distributed through spray nozzles at A-header location, (45-foot elevation), the solids deposits were cleaned by a steam soot blower. During operation with slurry distributed through spray nozzles at B-header location (35-foot elevation), solids deposits formed on the scrubber walls and nozzle headers because there was no soot blower at this elevation. During several tests, large pieces of these deposits fell and damaged the grids below the spray header. This problem was initially solved by improving grid supports and increasing the depth of grids per stage from 1-1/4 inches to 3-3/4 inches, or by operating with a 45-foot tower. Problems with pluggage of the Bete spiral-tip spray nozzles and tower flooding have not been observed.

Later, a presaturator, two, 1-inch Bete spray nozzles, was installed at the top of the tower (A-header) immediately below the steam soot blower. Solids deposition and flow area restriction were reduced but not eliminated. To improve gas/liquor contact and slurry distribution in the presaturator, the original two spray nozzles were replaced with four, 3/4-inch Bete spray nozzles. Tests with these nozzles are in progress.

After the scrubber sump divider was extended and the 45-degree baffle plate was installed in the 90-degree bend upstream of the mist eliminator housing, increased solids deposition was observed in this sump region. A soot blower had previously cleaned this area but could not be operated after the installation of the 45-degree baffle plate.

Solids deposits at the wet-dry interface and in the scrubber outlet duct should not be a problem in full-scale operation if soot blowers are installed in these areas.

During startup a vibration was encountered in the mist eliminator or reheater section at air rates above approximately 27,000 aft^3/min (100°F). Because of the vibration, the maximum scrubber gas velocity tested was restricted to 26.7 ft/sec or an inlet flue gas rate of 25,000 aft^3/min (300°F).

Mist Eliminator Performance

Reliable operation of the mist eliminator was not achieved during the EPRI test program for superficial gas velocities through the scrubber and mist eliminator of 26.7 ft/sec and 18.5 ft/sec, respectively. None of the mist eliminator wash schemes or types of mist eliminators tested were successful. Apparently, the poor gas distribution at the outlet of the scrubber caused excessive levels of slurry to be entrained into the mist eliminator, resulting with the mist eliminator pluggage. (Attempts to measure the entrainment loading were unsuccessful.) Poor gas distribution entering the mist eliminator also caused the localized velocities in the mist eliminator to exceed its maximum recommended operating velocity. The excessive entrainment loading entering the mist eliminator and excessive velocities in the mist eliminator contributed to its poor efficiency which caused the reheater pluggage. After the superficial gas velocity through the mist eliminator was reduced from 18.5 to 14 ft/sec (26.7 to 20 ft/sec through the tower), solids deposits in the mist eliminators and the reheater were reduced to minor deposits which did not interfere with operation of the scrubber. An events log of the mist eliminator configurations and wash schemes tested is presented in Table XI.

Turning vanes are being installed in the 90-degree bend upstream of the mist eliminator housing and an open-vane mist eliminator is being installed in the scrubber outlet vertical duct. This is an attempt to improve the gas distribution leaving the scrubber sump and entering the mist eliminator and permit reliable system operation with higher velocities (27 ft/sec) in the tower. After the gas distribution problem is solved, the effects of lower alkali stoichiometries, scrubber internal design, and other operating conditions, such as recirculated slurry rate upon mist eliminator efficiency and reliability can be tested.

Reheater Performance

Although the performance of the inline, indirect steam reheater was restricted by steam leaks in its head flanges and tubes, and excessive solids deposits on the tubes (due to poor mist eliminator efficiency), operation of the reheater did not interfere with cocurrent scrubber operation.

Steam leaks and solids deposits in the reheater permitted only a qualitative evaluation of its heat transfer characteristics. During the initial air-water tests, the overall heat transfer coefficient for clean tubes was calculated to be 15-20 Btu/hr-ft^2-^{o}F. The calculated heat transfer coefficient for fouled tubes and a partially plugged reheater ranged from 14 to 18 Btu/hr-ft^2-^{o}F. The heat transfer coefficient increased with an increase in flue gas rate.

SUMMARY

The prototype cocurrent scrubber test program at Shawnee successfully demonstrated that the cocurrent scrubber is an effective SO_2 scrubber when operated as a grid tower. During the EPRI test program, while operating at a superficial gas velocity of 27 ft/sec, an SO_2 removal efficiency of greater than 90% and a particulate removal efficiency of approximately 99% were achieved with lime or limestone absorbent. The inlet SO_2 concentration and particulate loading were approximately 2,500 ppm and 3 gr/sft^3, respectively.

Total system reliability was not demonstrated at the 27 ft/sec scrubber gas velocity because of solids deposition in the mist eliminator and reheater. None of the scrubber operating conditions or mist eliminator configurations or wash schemes tested provided reliable operating conditions at this scrubber gas velocity.

The solids deposits were apparently caused by the poor distribution of gas leaving the scrubber and entering the mist eliminator. The poor gas distribution leaving the scrubber increased the amount of entrained slurry, above levels expected with good gas distribution. Also, the distorted gas distribution entering the mist eliminator unevenly distributed the entrained slurry into the mist eliminator, distorted the wash spray pattern, and caused high localized velocities in the mist eliminator which reentrained the slurry into the reheater. Significant solids deposits in the mist eliminator and reheater have been essentially eliminated during the TVA test program by operating the scrubber at a 20 ft/sec superficial gas velocity. Apparently, the improved reliability at the lower scrubber gas velocity is provided by decreased entrained slurry from the scrubber and lower localized gas velocities through the mist eliminator.

Emphasis in the continuing test program will be on obtaining reliable operation of the mist eliminator and reheater at a scrubber gas velocity of 27 ft/sec and testing of forced oxidation. Equipment modifications which are required for these tests are in progress.

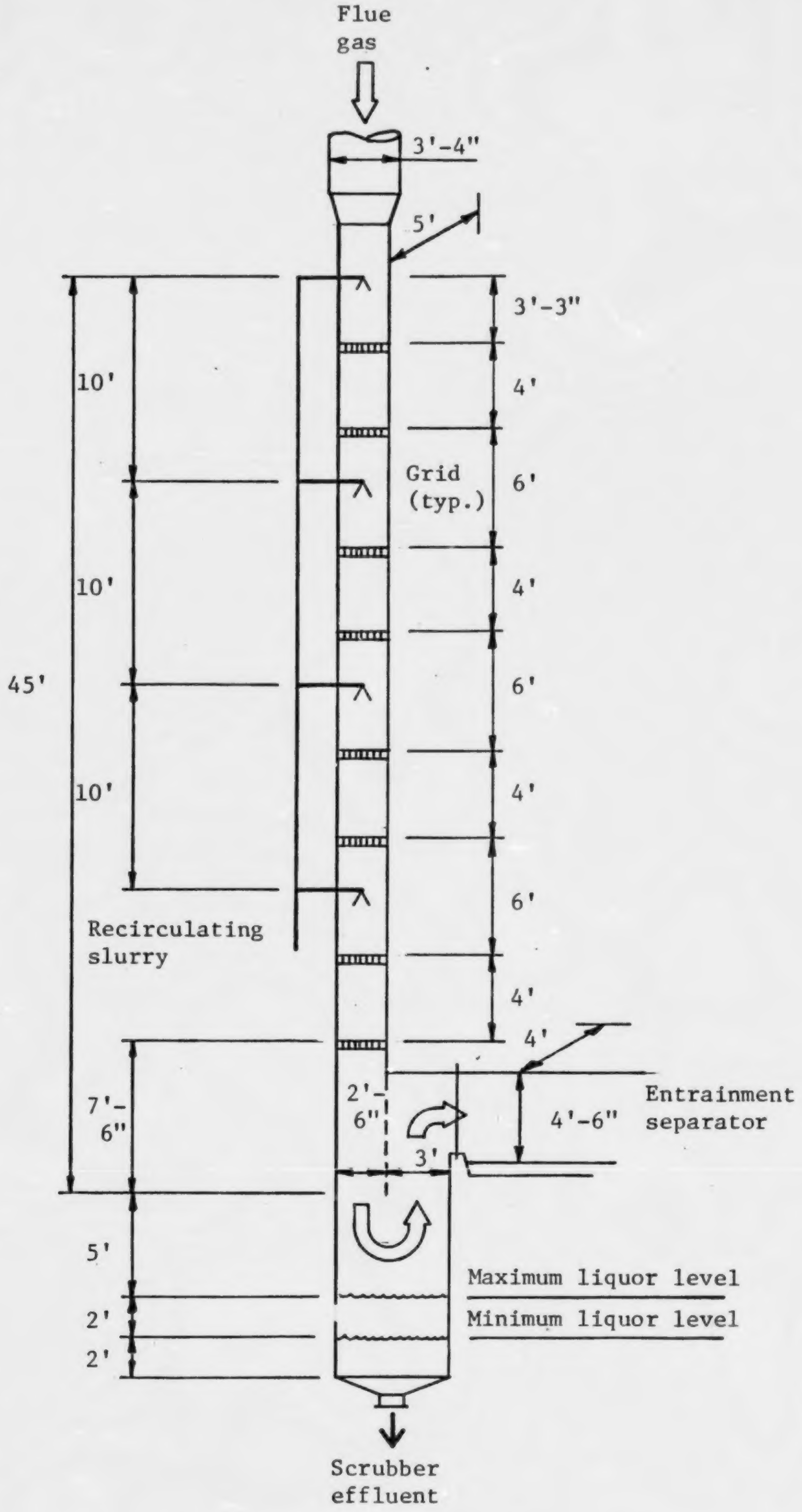

Flue gas
3'-4"
5'
3'-3"
10'
4'
Grid (typ.)
6'
10'
4'
45'
6'
10'
4'
6'
Recirculating slurry
4'
4'
7'-6"
2'-6"
4'-6"
Entrainment separator
3'
5'
Maximum liquor level
2'
Minimum liquor level
2'
Scrubber effluent

FIGURE 1

Cocurrent Scrubber

Schematic

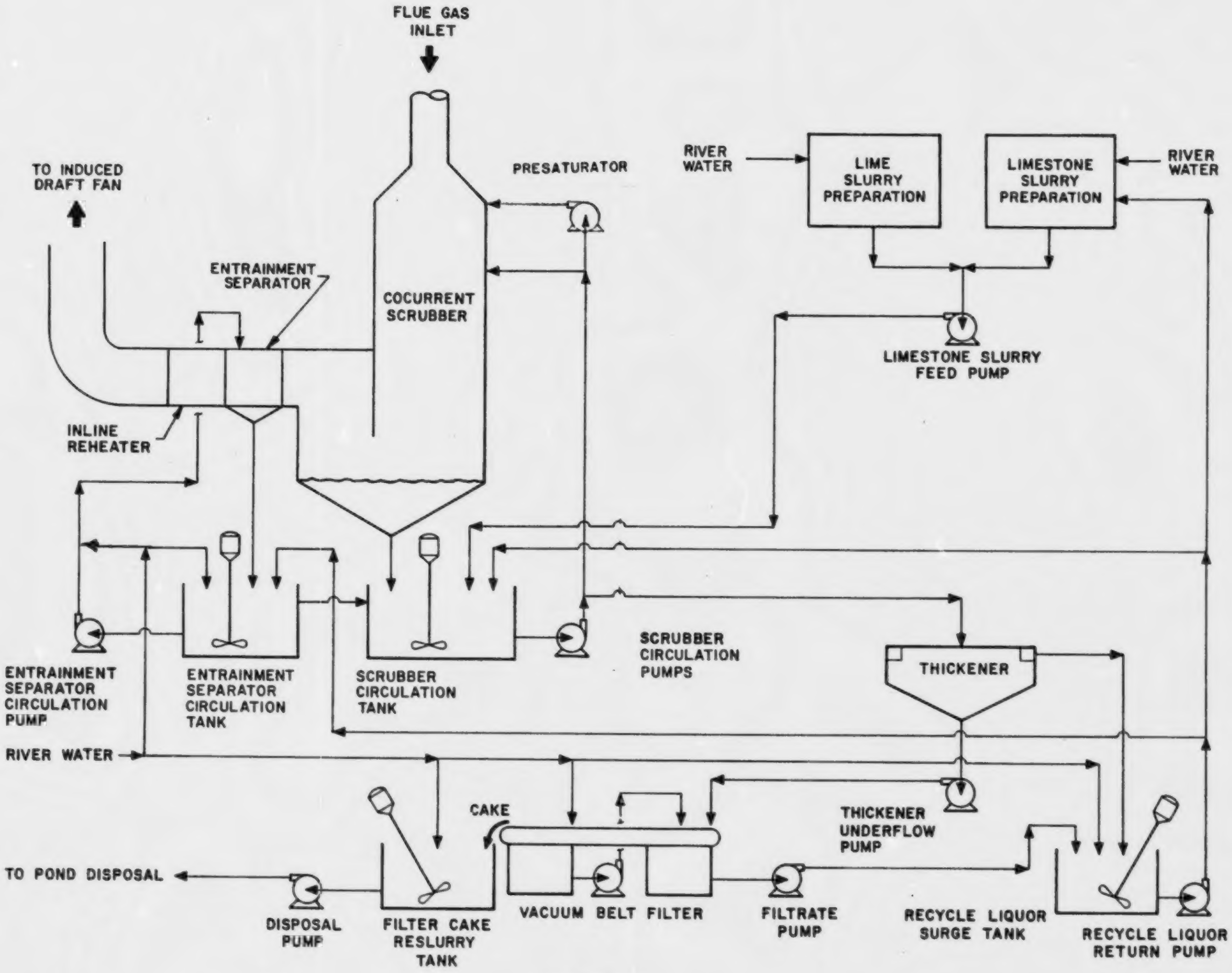

FLUE GAS INLET
TO INDUCED DRAFT FAN
PRESATURATOR
RIVER WATER
LIME SLURRY PREPARATION
LIMESTONE SLURRY PREPARATION
RIVER WATER
ENTRAINMENT SEPARATOR
COCURRENT SCRUBBER
LIMESTONE SLURRY FEED PUMP
INLINE REHEATER
SCRUBBER CIRCULATION PUMPS
THICKENER
ENTRAINMENT SEPARATOR CIRCULATION PUMP
ENTRAINMENT SEPARATOR CIRCULATION TANK
SCRUBBER CIRCULATION TANK
RIVER WATER
CAKE
THICKENER UNDERFLOW PUMP
TO POND DISPOSAL
DISPOSAL PUMP
FILTER CAKE RESLURRY TANK
VACUUM BELT FILTER
FILTRATE PUMP
RECYCLE LIQUOR SURGE TANK
RECYCLE LIQUOR RETURN PUMP

FIGURE 2

Cocurrent Scrubber

Shawnee Steam Plant Test Facility

Flow Diagram

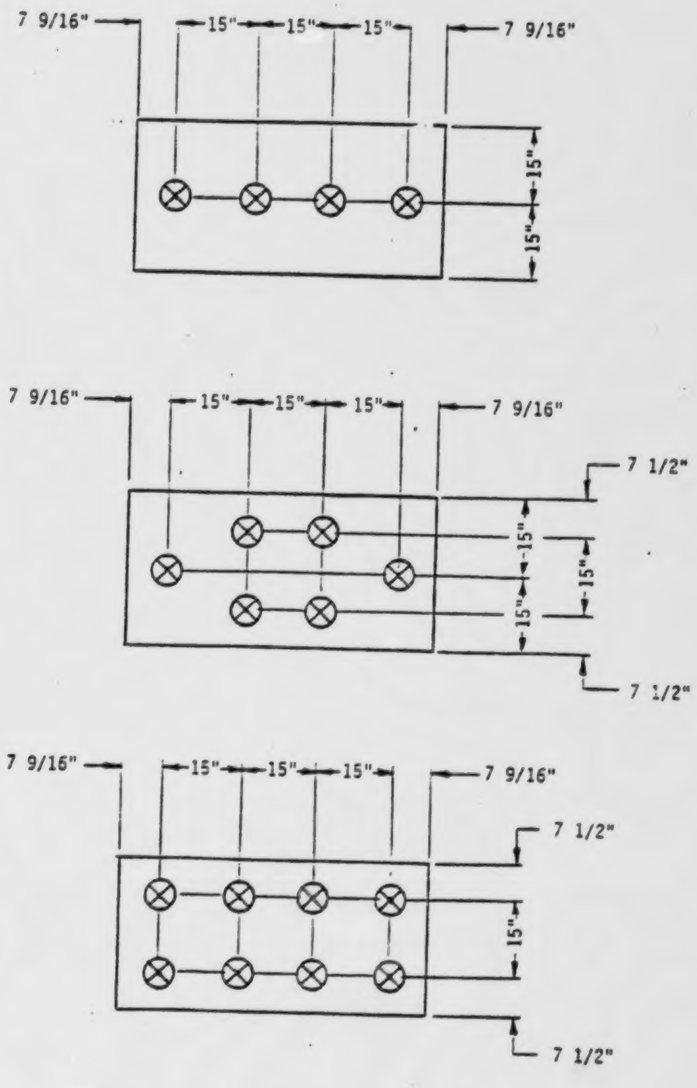
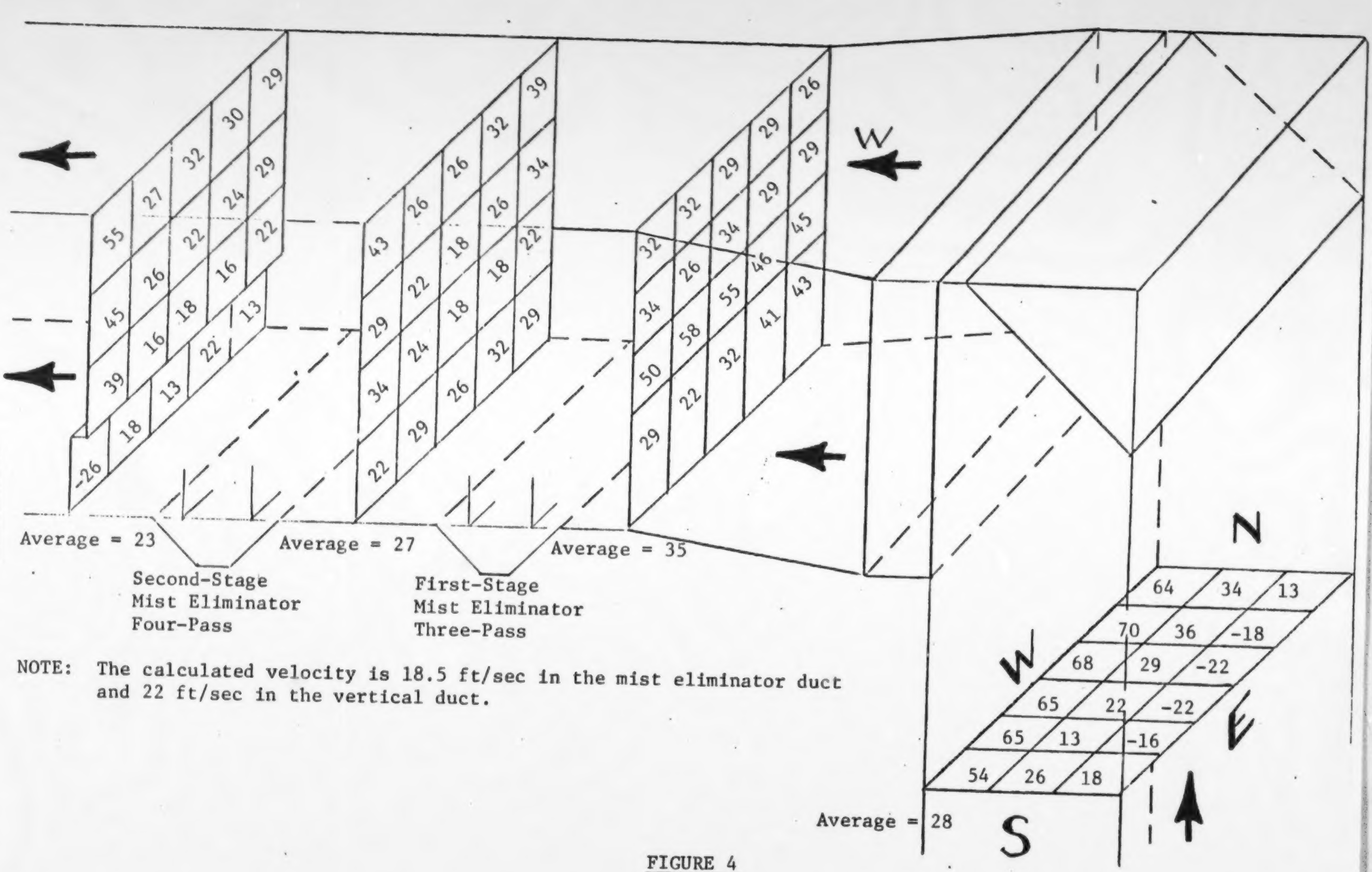

7 9/16"
15"
15"
15"
7 9/16"
15"
15"
7 9/16"
15"
15"
15"
7 9/16"
7 1/2"
15"
15"
15"
7 1/2"
7 9/16"
15"
15"
15"
7 9/16"
7 1/2"
15"
7 1/2"

FIGURE 3

Cocurrent Scrubber Cross-Section Presenting Nozzle

Locations for Four-, Six-, and Eight-Nozzle Header Configurations


W
W
N
E
S
55
27
32
30
29
45
26
22
24
29
39
16
18
16
22
-26
18
13
22
13
43
26
26
32
39
29
22
18
26
34
34
24
18
18
22
22
29
26
32
29
32
32
29
29
26
34
26
34
29
29
50
58
55
46
45
29
22
32
41
43
64
34
13
70
36
-18
68
29
-22
65
22
-22
65
13
-16
54
26
18
Average = 23
Average = 27
Average = 35
Average = 28
Second-Stage
Mist Eliminator
Four-Pass
First-Stage
Mist Eliminator
Three-Pass

NOTE: The calculated velocity is 18.5 ft/sec in the mist eliminator duct and 22 ft/sec in the vertical duct.

FIGURE 4

Cocurrent Scrubber Mist Eliminator Velocity Profile (ft/sec)

Air-Slurry Test D-40

TABLE I

Highlights of Sodium Carbonate Factorial Tests

Test Block No. 1

Cocurrent Scrubber

Test number	010[a]	130A[a]	100	110	120	130	140	150	160	170	180	190	191[b]
Scrubber operating conditions													
Physical configuration[c]													
Height, ft	45	45	45	45	45	45	45	45	45	45	45	45	45
Header location(s)[d]	A	A	A	A	A	A	A	A	A	A	A	A	A
Flue gas[e]													
Flowrate (inlet) aft^3/min	25,000	25,000	19,200	12,000	19,200	25,000	19,200	25,000	19,200	17,000	19,200	17,000	17,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	20.5	12.8	20.5	26.7	20.5	26.7	20.5	18.1	20.5	18.1	18.1
Slurry													
Recirculation rate, gpm	2,400	2,400	1,700	1,435	1,700	2,400	1,700	1,435	1,700	2,400	1,700	835	835
L/G, gal/mcf	120	120	111	149	111	120	111	72	111	176	111	61	61
Laboratory results of recirculated solution													
Composition, ppm													
Na	12,300	10,125	11,400	11,100	10,700	10,000	10,700	10,200	9,700	10,300	10,900	10,500	-
Sulfite (as SO_3)	-	6,785	7,237	8,368	7,011	4,071	6,559	7,690	4,071	4,976	2,714	9,047	-
Total dissolved solids	-	-	30,885	-	-	-	28,521	30,604	-	38,800	41,506	-	-
pH	-	7.13	7.87	8.01	8.11	7.99	7.87	8.07	7.94	8.10	8.14	8.01	-
Stoichiometry, mols Na/mol SO_2 absorbed	-	2.05	2.92	2.72	3.30	2.72	2.70	2.97	2.67	3.73	3.54	2.91	-
SO_2 removal efficiency, %	97	79	97	99	97	97	97	93	97	98	96	96	96

---

a. Startup runs to bring system to equilibrium.
b. Stoichiometric control instead of constant pH control which was used for test block No. 1.
c. No grids.
d. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
e. "Fly ash-free" flue gas.

TABLE II

Highlights of Sodium Carbonate Factorial Tests

Test Block No. 2

Cocurrent Scrubber

Test number	210	220	230	240	250	260	270	280	290	200
Scrubber operating conditions										
Physcial configuration[a]										
Height, ft	25	25	25	25	25	25	25	25	25	25
Header location(s)[b]	C	C	C	C	C	C	C	C	C	C
Flue gas[c]										
Flowrate (inlet) aft^3/min	19,200	12,000	19,200	25,000	19,200	25,000	19,200	17,000	19,200	17,000
Superficial velocity at 125°F, ft/sec	20.5	12.8	20.5	26.7	20.5	26.7	20.5	18.1	20.5	18.1
Slurry										
Recirculation rate, gpm	1,700	1,435	1,700	2,400	1,700	1,435	1,700	2,400	1,700	835
L/G, gal/mcf	111	149	111	120	111	72	111	176	111	61
Laboratory results of recirculated solution										
Composition, ppm										
Na	10,400	10,100	10,300	9,800	9,300	9,500	11,000	11,000	10,500	9,500
Sulfite (as SO_3)	12,439	10,178	11,987	8,821	9,499	11,308	11,761	5,880	7,690	2,262
Total dissolved solids	30,830	29,761	28,000	31,110	27,005	30,000	31,000	34,773	39,200	26,882
pH	7.50	7.40	7.15	7.49	7.10	7.10	7.36	7.58	6.96	7.30
Stoichiometry, mols Na/mol SO_2 absorbed	2.21	2.39	2.32	2.13	2.08	2.10	2.34	2.55	2.29	2.38
SO_2 removal efficiency, %	94	96	92	92	91	84	92	97	91	85

---

a. No grids.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. "Fly ash-free" flue gas.

TABLE III

Highlights of Sodium Carbonate Factorial Tests

Test Block No. 3

Cocurrent Scrubber

Test number	300	310	320	330	340	350	360	370
Scrubber operating conditions								
Physical configuration[a]								
Height, ft	25	25	25	25	25	25	25	25
Header location(s)[b]	C,D	C,D	C,D	C,D	C,D	C,D	C,D	C,D
Flue gas[c]								
Flowrate (inlet) aft^3/min	25,000	17,000	19,200	12,000	19,200	25,000	19,200	17,000
Superficial velocity at 125°F, ft/sec	26.7	18.1	20.5	12.8	20.5	26.7	20.5	18.1
Slurry								
Recirculation rate, gpm	2,400	2,400	1,700	1,435	1,700	1,435	1,700	835
L/G, gal/mcf	120	176	111	149	111	72	111	61
Laboratory results of recirculated solution								
Composition, ppm								
Na	10,300	10,700	10,400	8,900	9,500	10,900	9,200	9,300
Sulfite (as SO_3)	7,916	6,333	7,011	7,464	8,821	13,118	11,535	10,630
Total dissolved solids	-	33,600	32,322	28,541	30,400	30,599	28,258	29,603
pH	7.10	7.20	7.10	7.42	7.56	7.54	7.13	6.76
Stoichiometry, mols Na/mol SO_2 absorbed	2.34	2.31	2.17	2.28	2.41	2.24	1.98	1.78
SO_2 removal efficiency, %	93	96	97	98	98	92	97	90

---

a. No grids.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. "Fly ash-free" flue gas.

TABLE IV

Highlights of Sodium Carbonate Factorial Tests

Test Block No. 4

Cocurrent Scrubber

Test number	400	410	420	430	431	432	433	440	441	442	450	460	470
Scrubber operating conditions													
Physical configuration[a]													
Height, ft	45	45	45	45	45	45	45	45	45	45	45	45	45
Header location(s)[b]	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D
Flue gas[c]													
Flowrate (inlet) aft^3/min	19,200	12,000	19,200	25,000	25,000	12,500	25,000	19,200	19,200	19,200	25,000	17,000	17,000
Superficial velocity at 125°F, ft/sec	20.5	12.8	20.5	26.7	26.7	13.3	26.7	20.5	20.5	20.5	26.7	18.1	18.1
Slurry													
Recirculation rate, gpm	1,700	1,435	1,700	1,435	1,000	1,000	2,000	1,700	1,000	1,435	2,400	2,400	835
L/G, gal/mcf	111	149	111	72	50	100	100	111	65	93	120	176	61
Laboratory results of recirculated solution													
Composition, ppm													
Na	10,000	10,500	9,500	10,100	9,000	9,700	11,300	10,100	9,700	9,400	10,300	10,200	10,000
Sulfite (as SO_3)	8,142	10,856	11,082	12,892	14,475	14,023	11,308	11,761	11,308	10,404	6,333	11,082	5,654
Total dissolved solids	33,600	29,595	29,600	31,477	27,090	29,600	32,096	30,977	28,771	30,479	27,000	33,197	26,000
pH	6.52	6.50	6.58	6.52	6.77	7.02	6.90	6.77	6.81	6.55	6.92	7.10	7.02
Stoichiometry, mols Na/mol SO_2 absorbed	1.88	2.13	1.74	1.82	1.70	1.95	2.10	1.94	1.81	1.69	2.02	1.97	2.24
SO_2 removal efficiency, %	98	99	98	95	91	98	98	98	94	98	98	99	97

---

a. No grids.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. "Fly ash-free" flue gas.

TABLE V-A

Highlights of Lime Factorial Tests

Cocurrent Scrubber

Test number	400	401	1000	1001	100	1300	300	301	302	200
Scrubber operating conditions										
Physical configuration										
Height, ft	45	45	45	45	45	45	45	45	45	35
Number of grids[a]	0	8	8	8	8	8	0	8	8	6
Header locations(s)[b]	A	A	A	A	A,B	A,B	A,B,C	A,B,C	A,B,C	B,C
Flue gas[c]										
Flowrate (inlet), aft^3/min	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	26.7	26.7	26.7	26.7	26.7	26.7	26.7	26.7
Slurry										
Recirculation rate, gpm	2,400	2,400	800	800	1,600	2,000	2,400	2,400	2,400	1,600
L/G, gal/mcf	120	120	40	40	80	100	120	120	120	80
Laboratory results										
Recirculated slurry										
Slurry										
Solids concentration, wt %	12.9	11.2	9.9	10.5	10.2	10.4	18.3	10.4	-	13.3
pH	8.60	8.03	-	10.10	8.71	6.43	8.10	7.60	-	8.70
Solids										
Stoichiometry, mols Ca/mol inlet SO_2	1.10	1.12	0.69	0.85	0.99	1.03	1.01	1.01	0.89	1.05
Liquor composition, ppm										
Sulfite (as SO_3)	136	226	-	136	158	136	204	136	-	68
Total dissolved solids	4,800	2,737	-	4,110	2,463	2,838	4,097	3,970	-	4,150
Thickener underflow										
Solids concentration, %	28.3	28.5	26.2	26.3	19.9	29.4	20.1	23.9	26.8	18.2
SO_2 removal efficiency, %	81[d]	94	59	69	84	88	79	85	76[e]	80[f]

---

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. Flue gas with full loading of fly ash.
d. Obtained 74% SO_2 removal at 0.94 stoichiometry (mols Ca/mol inlet SO_2).
e. Obtained 74% SO_2 removal at 0.78 stoichiometry (mols Ca/mol inlet SO_2).
f. Obtained 76% SO_2 removal at 0.99 stoichiometry (mols Ca/mol inlet SO_2).

TABLE V-B

Highlights of Lime Factorial Tests

Cocurrent Scrubber

Test number	500	800	600	900	1400A	1400C	1400B[a]	1401	1500
Scrubber operating conditions									
Physical configuration									
Height, ft	35	35	25	25	35	35	35	35	35
Number of grids[b]	6	6	4	4	6	6	6	6	6
Header location(s)[c]	B	B	C	C	B	B	B	B	B
Flue gas[d]									
Flowrate (inlet), aft^3/min	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	26.7	26.7	26.7	26.7	26.7	26.7	26.7
Slurry									
Recirculation rate, gpm	2,400	1,600	2,400	1,600	2,000	2,400	2,400	2,400	1,600
L/G, gal/mcf	120	80	120	80	100	120	120	120	80
Laboratory results									
Recirculated slurry									
Slurry									
Solids concentration, wt %	10.0	11.4	10.4	10.8	11.0	11.7	10.2	10.3	9.4
pH	7.55	8.41	8.40	10.50	9.14	9.02	8.30	9.06	9.80
Solids									
Stoichiometry, mols Ca/mol inlet SO_2	1.03	1.01	1.02	1.02	1.00	0.99	0.96	0.97	1.02
Liquor composition, ppm									
Sulfite (as SO_3)	113	158	181	136	113	158	204	23	158
Total dissolved solids	4,080	3,152	4,357	4,363	4,000	2,692	4,453	4,576	1,694
Thickener underflow									
Solids concentration, %	26.4	26.7	–[e]	–[e]	–[e]	–[e]	–[e]	–[e]	–[e]
SO_2 removal efficiency, %	91	82	85	81	83	80	73	82	75

---

a. Constant pH control at 8.0 ± 0.3 instead of stoichiometric control.
b. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.
c. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; D, 387 ft.
d. Flue gas with full loading of fly ash.
e. Solids disposal system out of order because of freezing problems.

TABLE VI

Highlights of Lime Short-Term Reliability Tests

Cocurrent Scrubber

Test No.	L-2000 L-2001 L-2002	L-2100	L-2200	L-2300	L-2400	L-2500
Scrubber operating conditions						
Physical configuration						
Height, ft	35	45	45	45	45	45
No. of grids[a]	6	8	8	8	8	8
Header location(s)[b]	B	A	A	A	A	A
Pressure drop	2.0-3.0	3.0-3.6	3.2-3.9	3.5-4.2	2.8-5.0	3.2-5.0
Flue gas[c]						
Flow rate (inlet), aft^3/min	25,000	25,000	25,000	25,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	26.7	26.7	26.7	26.7
Inlet SO_2 concentration	1,350-2,900	1,500-3,000	1,600-3,200	2,200-3,200	2,750-3,200	2,500-3,200
Slurry						
Recirculation rate, gpm	2,400	2,400	2,400	2,400	1,200	1,800
L/G, gal/kft^3	120	120	120	120	60	90
Laboratory results						
Recirculated slurry						
Slurry						
Solids concentration, wt %	8.7-13.3	10.6-12.4	13.5-16.2	14.5-16.0	13.2-17.4	14.3-16.4
pH	5.8-9.2	6.2-8.0	5.8-8.6	7.5-8.8	10.4-11.0	8.3-9.3
Solids						
Stoichiometry, mols Ca/mol inlet SO_2	0.82-1.21	0.97-1.03	0.96-1.12	1.06-1.14	0.95-1.10	1.00-1.10
Liquor composition, ppm						
Sulfite (as SO_3)	23-339	113-181	90-249	362-407	68-181	90-181
Total dissolved solids	2,752-5,875	3,400-5,070	2,907-8,764	6,164-7,946	1,007-5,600	1,280-2,928
Thickener underflow[d]						
Solids concentration, %	27.0-31.5	22.2-48.6	23.3-34.8	33.3-33.9	20.6-30.4	23.8-27.3
SO_2 removal efficiency, %	75-95	89-92	90-97	97-99	84-93	88-94

---

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.

b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.

c. Flue gas with full loading of fly ash.

d. Thickener was not in operation for tests L-2000 and L-2001.

TABLE VII-A

Highlights of Limestone Factorial Tests

Cocurrent Scrubber

Test number	1200	1301	1300	100	200	1700	1600
Scrubber operating conditions							
Physical configuration							
Height, ft	45	45	45	45	45	45	45
Number of grids[a]	0	0	0	8	8	8	8
Header location(s)[b]	A	A	A	A	A	A	A
Flue gas[c]							
Flowrate (inlet), aft^3/min	25,000	25,000	20,000	25,000	12,000	25,000	12,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	21.3	26.7	12.8	26.7	12.8
Slurry							
Recirculation rate, gpm	2,400	2,400	2,400	2,400	1,000	1,000	2,400
L/G, gal/mcf	120	120	150	120	120	50	250
Laboratory results							
Recirculated slurry							
Slurry							
Solids concentration, wt %	15.5	15.7	13.7	15.8	16.1	15.8	16.1
pH	6.45	6.25	6.21	5.97	6.05	5.89	5.40
Solids							
Stoichiometry, mols Ca/mol inlet SO_2	1.30	1.10	1.22	1.29	1.33	1.09	1.07
Liquor composition, ppm							
Sulfite (as SO_3)	136	226	158	158	294	317	633
Total dissolved solids	5,172	4,849	4,010	8,000	7,579	8,156	7,686
Thickener underflow							
Solids concentration, %	29.3	32.6	29.5	25.8	33.4	32.0	33.2
SO_2 removal efficiency, %	71	69[d]	82[e]	94	95	75	96[f]

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. Flue gas with full loading of fly ash.
d. Obtained 71% SO_2 removal at 1.27 stoichiometry (mols Ca/mol inlet SO_2).
e. Obtained 77% SO_2 removal at 1.02 stoichiometry (mols Ca/mol inlet SO_2).
f. Obtained 92% SO_2 removal at 0.94 stoichiometry (mols Ca/mol inlet SO_2).

TABLE VII-B

Highlights of Limestone Factorial Tests

Cocurrent Scrubber

Test number	300	300B	301	500	501	502	400	503	1400	1500
Scrubber operating conditions										
Physical configuration										
Height, ft	45	45	45	45	45	45	45	45	45	45
Number of grids[a]	0	0	0	8	8	8	8	8	8	8
Header location(s)[b]	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D	A,B,C,D
Flue Gas[c]										
Flowrate (inlet), aft^3/min	25,000	25,000	12,000	25,000	25,000	20,000	12,000	20,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	12.8	26.7	26.7	21.3	12.8	21.3	26.7	12.8
Slurry										
Recirculation rate, gpm	2,400	2,400	2,400	1,000	1,500	1,500	2,400	1,500	2,400	1,000
L/G, gal/mcf	120	120	250	50	75	94	250	94	120	104
Laboratory results										
Recirculated slurry										
Slurry										
Solids concentration, wt %	16.3	13.8	-	16.5	16.4	16.4	14.3	15.8	15.2	14.4
pH	6.11	6.26	-	6.22	6.33	6.25	5.94	6.06	5.83	5.99
Solids										
Stoichiometry, mols Ca/mol inlet SO_2	1.39	1.33	-	1.53	1.35	1.32	1.58	1.15	0.99	1.02
Liquor composition, ppm										
Sulfite (as SO_3)	271	181	-	90	158	136	204	158	158	204
Total dissolved solids	6,000	4,400	-	5,714	4,545	5,721	4,400	7,350	4,400	7,089
Thickener underflow										
Solids concentration, %	30.2	25.5	-	40.0	28.2	29.0	28.2	28.5	34.5	37.0
SO_2 removal efficiency, %	80	77	99	65	78	85	99	82[d]	78	81

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. Flue gas with full loading of fly ash.
d. Obtained 80% SO_2 removal at 1.03 stoichiometry (mols Ca/mol inlet SO_2).

TABLE VII-C

Highlights of Limestone Factorial Tests

Cocurrent Scrubber

Test number	1100	1000	901	800	900	600	700
Scrubber operating conditions							
Physical configuration							
Height, ft	35	35	25	25	25	25	25
Number of grids[a]	6	6	4	4	4	4	4
Header location(s)[b]	B	B	C	C	C	C,D	C,D
Flue gas[c]							
Flowrate (inlet), aft^3/min	25,000	25,000	25,000	25,000	12,000	25,000	12,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	26.7	26.7	12.8	26.7	12.8
Slurry							
Recirculation rate, gpm	1,000	2,400	1,000	1,000	2,400	2,400	1,000
L/G, gal/mcf	50	120	50	50	250	120	104
Laboratory results							
Recirculated slurry							
Slurry							
Solids concentration, wt %	19.2	15.6	-	14.5	15.4	15.5	15.3
pH	6.19	6.16	-	6.08	5.77	6.22	6.07
Solids							
Stoichiometry, mols Ca/mol inlet SO_2	1.33	1.31	0.66	1.32	1.37	1.30	1.21
Liquor composition, ppm							
Sulfite (as SO_3)	136	249	-	90	204	90	158
Total dissolved solids	4,571	5,728	-	8,000	4,000	7,342	6,000
Thickener underflow							
Solids concentration, %	27.3	31.9	36.7	32.6	33.8	30.4	33.4
SO_2 removal efficiency, %	71	92	47	64	96	70	78

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.
b. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.
c. Flue gas with full loading of fly ash.

TABLE VIII-A

Highlights of Limestone Short-Term Reliability Tests

Cocurrent Scrubber

Test No.	LS-2100[a],[b]	LS-2200[b]	LS-2300[b]	LS-2400	LS-2201	LS-2900
On-stream time, hr	76	157	161	158	38	58
Scrubber operating conditions						
Physical configuration						
Height, ft	45	45	45	35	45	45
Number of grids[c],[d]	8	8	8	6	8	0
Header location[e]	A	A	A	B	A	A
Pressure, drop, in. H_2O	3.2-3.8	2.6-3.6	2.5-3.1	2.4-3.5	2.8-3.6	0.1
Flue gas[f]						
Flow rate (inlet), aft^3/min	25,000	25,000	25,000	25,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7	26.7	26.7	26.7	26.7
Inlet SO_2 concentration, ppm	2,320-2,720	2,340-2,800	2,160-3,120	1,840-3,000	2,280-2,800	1,920-2,480
Slurry[g]						
Recirculation rate, gpm	2,400	2,400	2,000	2,400	2,400	2,400
L/G, gal/kft^3	120	120	100	120	120	120
Laboratory results						
Recirculated slurry						
Solids concentration, wt %	14.5-16.2	14.9-16.3	14.0-16.1	14.8-17.2	14.1-15.7	13.7-15.5
pH	6.1-6.3	6.0-6.3	6.1-6.4	6.0-6.3	5.9-6.2	6.1-6.3
Total dissolved solids, ppm	2,536-3,873	2,467-5,200	2,224-5,200	2,653-8,330	7,812-9,938	7,099-9,917
Solids						
Stoichiometry, mols Ca/mol inlet SO_2	1.25-1.38	1.20-1.43	1.21-1.44	1.25-1.40	1.19-1.42	1.05-1.51
Thickener underflow						
Solids concentration, wt %	20.9-25.9	20.0-30.9	17.7-29.6	18.0-27.7	19.8-22.7	19.8-26.5
Filter cake						
Solids concentration, wt %	-	-	-	59.6-64.4	60.6-65.4	71.7-70.9
SO_2 removal efficiency, %	93-95	93-95	89-91	88-91	91-95	77-81

---

a. Used first-stage Heilex mist eliminator only and distributed slurry in the tower through 90-degree spray nozzles. Remaining tests used two-stage Heilex mist eliminator and 60-degree spray nozzles.

b. Horizontal-belt filter was not installed for these tests.

c. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.

d. Depth of grids was 1-1/4 inches/elevation.

e. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft;, and D, 387 ft.

f. Flue gas with full loading of fly ash.

g. Recirculated slurry distributed through 90-degree spray nozzles for test LS-2100. For the other tests, 60-degree spray nozzles were used.

TABLE VIII-B

Highlights of Limestone Short-Term Reliability Tests

Cocurrent Scrubber

Test No.	LS-2901	LS-2700	LS-2701	LS-2702	LS-2703	LS-3000
On-stream time, hr	96	18	143	93	296	80
Scrubber operating conditions						
Physical configuration						
Height, ft	45	35	35	35	35	35
Number of grids[a,b]	0	6	6	6	6	6
Header location[c]	A	B	B	B	B	B
Pressure drop, in. H_2O	-0.6	6.7-7.2	6.8-7.7	6.3-7.5	6.0-7.1	6.9-8.1
Flue gas[d]						
Flow rate (inlet), aft^3/min	17,000	25,000	25,000	25,000	25,000	25,000
Superficial velocity at 125°F, ft/sec	18.1	26.7	26.7	26.7	26.7	26.7
Inlet SO_2 concentration, ppm	2,000-2,720	2,320-2,640	2,320-3,080	2,080-2,600	2,000-2,720	2,400-2,800
Slurry						
Recirculation rate, gpm	2,400	1,800-2,400	1,600-1,800	1,400	1,400	1,400-1,800
L/G, gal/kft^3	176	90-120	80-90	70	70	70-90
Laboratory results						
Recirculated slurry						
Solids concentration, wt %	14.2-16.7	13.0-14.5	14.7-16.3	13.4-16.6	12.9-18.7	14.5-16.7
pH	6.0-6.3	5.9-6.1	5.9-6.4	5.7-6.2	5.9-6.4	5.9-6.2
Total dissolved solids, ppm	3,940-12,580	7,540-9,070	4,280-9,620	7,120-9,210	3,110-8,190	2,135-7,688
Solids						
Stoichiometry, mols Ca/mol inlet SO_2	1.07-1.37	1.07-1.39	1.10-1.48	1.17-1.52	1.16-1.47	1.03-1.30
Thickener underflow						
Solids concentration, wt %	18.9-31.0	23.4-27.3	19.0-26.6	18.2-24.0	16.8-28.2	21.6-31.3
Filter cake						
Solids concentration, wt %	59.9-70.6	68.8-68.9	58.1-79.8	69.0-73.3	60.9-85.2	68.8-75.6
SO_2 removal efficiency, %	87-92	88-92	89-94	87-94	87-92	86-90

---

a. Grid elevations: 412, 408, 402, 398, 392, 388, 382, and 378 ft.

b. Depth of grids was 3-3/4 inches/elevation.

c. Header elevations: A, 417 ft; B, 407 ft; C, 397 ft; and D, 387 ft.

d. Flue gas with full loading of fly ash.

TABLE IX

Highlights of Limestone Long-Term Reliability Tests

Cocurrent Scrubber

Test No.	LS-4000	LS-4001
On-stream time, hr	349	182
Scrubber operating conditions		
Physical configuration		
Height, ft	38	38
Number of grids[a,b]	6	6
Header location[c]	B	B
Pressure drop, in. H_2O	7.5-9.0	8.0-12.0
Flue gas[d]		
Flow rate (inlet), aft^3/min	25,000	25,000
Superficial velocity at 125°F, ft/sec	26.7	26.7
Inlet SO_2 concentration, ppm	1,500-3,000	1,100-2,500
Slurry		
Recirculation rate, gpm	1,600-1,800	1,800-2,400
L/G, gal/kft^3	80-90	90-120
Laboratory results		
Recirculated slurry		
Solids concentration, wt %	13.8-16.3	14.33-16.0
pH	6.0-6.3	5.6-6.3
Total dissolved solids, ppm	3,200-12,700	5,700-14,000
Solids		
Stoichiometry, mols Ca/mol inlet SO_2	1.09-1.48	1.03-1.22
Thickener underflow		
Solids concentration, wt %	20.4-30.0	21.7-36.6
Filter cake		
Solids concentration, wt %	56.1-64.3	59.9-69.4
SO_2 removal efficiency, %	81-94	70-97

---

a. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.
b. Depth of grids was 3-3/4 inches/elevation.
c. Header elevation: B, 407 ft.
d. Flue gas with full loading of fly ash.

TABLE X

Highlights of Limestone Mist Eliminator Reliability Tests

Cocurrent Scrubber

Test No.	LS-5000C	LS-5001C	LS-5002C	LS-5010C
On-stream time, hr	319	470	348	238
Scrubber operating conditions				
Physical configuration				
Height, ft	38	38	38	38
Number of grids[a,b]	6	6	6	6
Header location[c]	B	B	B	B
Pressure drop, in. H_2O	3.5-4.0	3.3-4.3	3.3-4.3	6.6-7.9
Flue gas[d]				
Flow rate (inlet), aft^3/min	18,750	18,750	18,750	25,000
Superficial velocity at 125°F, ft/sec	20.0	20.0	20.0	27.0
Inlet SO_2 concentration, ppm	1,340-2,120	1,320-2,320	1,840-2,440	1,360-2,680
Slurry[e]				
Recirculation rate, gpm	1,400	1,495	1,495	1,895
L/G, gal/kft^3	93	100	100	95
Laboratory results				
Recirculated slurry				
Solids concentration, wt %	8.3-11.5	8.2-11.5	8.7-11.4	8.7-11.4
pH	6.0-6.4	5.93-6.43	5.83-6.40	5.85-6.29
Total dissolved solids, ppm	4,100-9,900	2,440-11,295	7,618-17,708	9,423-15,144
Solids				
Stoichiometry, mols Ca/mol inlet SO_2	1.12-1.46	1.16-1.52	1.18-1.52	1.16-1.47
Thickener underflow				
Solids concentration, wt %	18.8-27.4	16.1-23.0	18.1-27.4	13.6-21.9
Filter cake				
Solids concentration, wt %	54.9-77.2	54.7-78.7	56.5-67.3	54.7-62.5
SO_2 removal efficiency, %	88-94	92-94	92-94	93-97

---

a. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.
b. Depth of grids was 3-3/4 inches/elevation.
c. Header elevation: B, 407 ft.
d. Flue gas with low loading of fly ash.
e. Includes presaturator rates.

TABLE XI

Mist Eliminator Wash Configurations

Cocurrent Scrubber

Date			First stage												Second stage					
				Wash on the upstream face of the first stage					Wash on the back side of the first stage						Wash on the upstream face of the second stage				
							Nozzle					Nozzle						Nozzle	
From	To	Tests	Type	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig	Type	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig
10/10/78	12/1/78	Sodium carbonate factorial	Heil, 4-pass FRP	Raw water	24	2/30	1/2	35	-	-	-	-	-	-	-	-	-	-	-
11/8/78	12/1/78	Limestone factorial	Heil, 4-pass FRP	Raw water	24	2/15	1/2	35	-	-	-	-	-	-	-	-	-	-	-
12/7/78	12/7/78	-	Heil, 4-pass Noryl thermoplastic	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12/8/78	1/11/79	Limestone and lime factorial	Munters 2-pass polypropylene	Raw water	24	2/15	1/2	35	-	-	-	-	-	Munters 2-pass polypropylene	Raw water	24	2/30	1/2	35
1/11/79	1/16/79	Lime factorial	Humboldt-Lamellar	Raw water	24	2/15	1/2	35	-	-	-	-	-	-	-	-	-	-	-
1/16/79	1/26/79	Lime factorial and reliability	-	-	-	-	-	-	-	-	-	-	-	Humboldt-Lamellar	Raw water	24	2/15	1/2	35
1/27/79	2/6/79	Lime reliability	-	-	-	-	-	-	-	-	-	-	-	Humboldt-Lamellar	Raw water	24	Continuous	1/2	35
2/6/79	2/13/79	Lime reliability	-	-	-	-	-	-	-	-	-	-	-	Humboldt-Lamellar	Raw water	24	2/15	1/2	35
2/13/79	2/27/79	Lime and limestone reliability	Heil, 4-pass 316 SS	Raw water	24	2/15	1/2	35	-	-	-	-	-	-	-	-	-	-	-
3/5/79	3/19/79	Limestone reliability	Heil, 4-pass 316 SS	Raw water	24	2/15	1/2	35	Raw water	24	2/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	2/30	3/4	7
3/20/79	3/26/79	Limestone reliability	Heil, 4-pass 316 SS	Raw water	24	Continuous	1/2	35	Raw water	24	2/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	2/30	3/4	7

TABLE XI (continued)

			First stage											Second stage					
				Wash on the upstream face of the first stage					Wash on the back side of the first stage						Wash on the upstream face of the second stage				
Date							Nozzle					Nozzle						Nozzle	
From	To	Tests	Type	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig	Type	Medium	Rate, gpm	Frequency, min. wash/interval	Size, in.	Pressure, psig
3/26/79	3/30/79	Limestone reliability	Heil, 4-pass 316 SS	Raw water recycle around D-304	24	Continuous	1/2	35	Raw water	24	2/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	2/30	3/4	7
3/30/79	4/6/79	Limestone reliability	Heil, 4-pass 316 SS	Raw water	24	Continuous	1/2	35	Raw water	24	2/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	2/30	3/4	7
4/6/79	4/10/79	Limestone reliability	Heil, 4-pass 316 SS	Recycle liquor and raw water	24 16	5/16 1/16	1/2	35 14	Raw water	24	1/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	1/30	3/4	7
4/10/79	4/16/79	Limestone reliability	Heil, 4-pass 316 SS	Recycle liquor and raw water	24 24	2.75/5 0.58/5	1/2	35	Raw water	24	1/30	3/4	7	Heil, 4-pass 316 SS	Raw water	24	1/30	3/4	7
4/17/79	4/30/79	Limestone reliability	Heil, 4-pass 316 SS	Recycle liquor and raw water	48 24	4.17/5 0.58/5	3/4	34 7	Raw water	24	2/30	1/2	35	Heil, 4-pass 316 SS	Raw water	24	2/30	1/2	35
5/1/79	5/4/79	Limestone reliability	Heil, 4-pass 316 SS	Recycle liquor and raw water	24	2.75/5 0.58/5	1/2	35	Raw water	24	1/30	1/2	35	Heil, 4-pass 316 SS	Raw water	24	1/30	1/2	35
7/5/79	7/31/79	Limestone reliability	Heil, 3-pass 316 SS	Recycle liquor and raw water	30	2.20/5 0.47/5	1/2	35	Raw water	30	0.8/30	1/2	35	Heil, 4-pass 316 SS	Raw water	30	0.8/30	1/2	35
8/8/79[a]	9/17/79	Limestone reliability	Heil, 3-pass 316 SS	Recycle liquor and raw water	30 30	1.67/5 0.33/5	1/2	35	Raw water	30	0.5/30	1/2	35	Heil, 4-pass 316 SS	Raw water	30	0.5/30	1/2	35
9/17/79	10/2/79	Limestone reliability	Heil, 3-pass 316 SS	Recycle liquor and raw water	30 30	0.83/4 0.17/4	1/2	35	Raw water	30	0.5/30	1/2	35	Heil, 4-pass 316 SS	Raw water	30	0.5/30	1/2	35
10/2/79	10/15/79	Limestone reliability	Heil, 3-pass 316 SS	Recycle liquor and raw water	30 30	2/5 0.47/5	1/2	35	Raw water	30	0.8/30	1/2	35	Heil, 4-pass 316 SS	Raw water	30	0.8/30	1/2	35

---

a. Solids deposition in the mist eliminator and reheater was insignificant for this scheme primarily due to the lower scrubber gas velocity (20 ft/sec). Solids deposition was a problem for all of the other schemes.

THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM

C. D. Stephenson and R. L. Torstrick
Emission Control Development Projects
Office of Power
Tennessee Valley Authority
Muscle Shoals, Alabama 35660

Prepared for Presentation at
Fifth Industry Briefing Conference
Results of EPA Lime/Limestone Wet Scrubbing Test Programs
Sponsored by the U.S. Environmental Protection Agency
Royal Villa Motel in Raleigh, North Carolina 27711
December 5, 1979

Preceding page blank

THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM

C. D. Stephenson and R. L. Torstrick
Emission Control Development Projects
Office of Power
Tennessee Valley Authority
Muscle Shoals, Alabama

ABSTRACT

In conjunction with the EPA-sponsored Shawnee test program, Bechtel National, Inc., and the Tennessee Valley Authority have jointly developed a computer program capable of projecting material balances and relative economics for lime and limestone scrubbing. In addition to a Turbulent Contact Absorber, a venturi - spray tower and a spray tower have been incorporated as scrubbing options. The program also has been expanded to include particulate removal costs and the use of chemical additives. The current program may also be used to project the effect of variations in lifetime operating profile, alternative sludge disposal methods, or additives (MgO or adipic acid) on costs for either process. Although the program is not intended to compute the economics of an individual system to a high degree of accuracy, it allows prospective users to quickly project comparative design and costs for various limestone and lime case variations on a common design and cost basis. The technical and economic bases for the current program are presented. Future process design alternatives to be incorporated into the program and limitations to the program validity are also discussed.

THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM

INTRODUCTION

The current status of flue gas desulfurization (FGD) in the United States is rapidly reaching the point where the capital investment for lime-limestone throwaway systems and their operating costs are fairly predictable. Much data are available from the installations in actual operation and others in planning or evaluation stages. The acceptance of limestone scrubbing as the most cost-effective FGD process for at least the short term is apparent. In light of the current FGD status, the utility industry is in need of a yardstick with which to evaluate the merits of alternate air pollution control plans. EPA, TVA, and Bechtel National, Inc., are attempting to provide a part of that yardstick with the development of the Shawnee Lime-Limestone Computer Economics program. This program is capable of projecting comparative investment and revenue requirements for lime and limestone scrubbing systems. The computer program has been developed to permit the rapid estimation of relative economics of these systems for variations in process design alternatives (i.e., limestone versus lime scrubbing, alternative scrubber types, or alternative sludge disposal methods) or variations in the values of independent design parameters (i.e., scrubber gas velocity and L/G ratio, alkali stoichiometry, slurry residence time, reheat temperature, and specific sludge disposal design). Although the program is not intended to compute the economics of an individual system to a high degree of accuracy, it is based on sufficient detail to allow the quick projection of preliminary conceptual design and costs for various lime-limestone case variations on a common design and cost basis.

PROGRAM DEVELOPMENT

The technical development of the Shawnee Lime-Limestone Computer Economics program is based on actual data obtained at the Shawnee Test Facility. Bechtel and TVA have shared the responsibility of the program development. Bechtel was responsible for analyzing the test results and developing the models which calculate the overall material balance flow rates and stream compositions. Bechtel provided these models to TVA, which in turn was responsible for determining the size limitations of the required equipment to establish the minimum number of parallel equipment trains, accumulating cost data for the major equipment items, and developing models for projecting equipment and field material costs as a function of equipment capacity. Utilizing these relationships TVA developed models to project the overall investment cost breakdown and a procedure for using the output of the material balance and investment models as input to a previously developed TVA model for projecting annual and lifetime revenue requirements.

PROGRAM DOCUMENTATION

TVA has presented three papers which report the status of the program development and displayed its capabilities. The first, titled "Shawnee Limestone-Lime Scrubbing Process Computerized Design-Cost Estimate Program: Summary Description Report," was given at the Industry Briefing Conference sponsored by EPA at Raleigh, North Carolina, October 14, 1976. The second, titled "Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization," was presented at the EPA Flue Gas Desulfurization Symposium in Hollywood, Florida, November 8-11, 1977. A third paper, titled "Current Status of Development of the Shawnee Lime-Limestone Computer Program," was presented at the EPA Industry Briefing Conference held in Raleigh, North Carolina, August 29, 1978.

Since August 1978 a number of additional options have been added to the program. These new program options include venturi - spray tower and spray tower scrubber alternatives, two forced oxidation methods, capability of using chemical additives, alternate upstream particulate removal devices, an SO_2 removal model relating SO_2 removal to input operating parameters and a 1979 equipment cost model update. This paper specifies the current program design basis and capabilities. TVA has prepared a users manual for the overall program which includes all the information required for running the Turbulent Contact Absorber (TCA) program option. Additional information is available from TVA detailing the inputs for running a newer version of the program which includes the options discussed in this paper.

NEW PROGRAM OPTIONS

To illustrate the current program capabilities an example run of the updated program is shown in the appendix. This example is for a venturi - spray tower system with forced oxidation and adipic acid addition. A discussion of the new options available in the program since August 1978 is given below.

Particulate Removal Model

A particulate removal model which predicts the installed cost of several particulate removal devices has been added to the program.

Venturi - Spray Tower

An option allowing the use of a venturi - spray tower is now available. This option can be used with one or two hold tanks. New inputs for this option require an L/G for the venturi and residence time for the venturi hold tank.

Spray Tower

Also available is the spray tower option. This option is essentially the same as the TCA option except for the difference in the design of the absorber unit.

SO_2 Removal

A new option has been added for all three absorber types that calculates the SO_2 removal for cases in which all of the operating parameters are defined. This is an alternate to the past SO_2 removal model in which percent removal or outlet emission was a required input.

Forced Oxidation

A forced oxidation option is now available for both in-loop and bleed-stream oxidation. A new input required for this option is the air stoichiometry, which in conjunction with the percent oxidation input allows the program to size the air supply equipment. At present, the forced oxidation options run on a force-through basis. This means that the user supplies the input for the amount of sulfite converted to sulfate. The models do not consider the effect of high sulfate concentrations on SO_2 removal; however, models which account for the effect of varying concentrations are planned for addition to the program.

Adipic Acid and Magnesium Oxide Addition

An option which allows for the sizing and costing of the equipment for adipic acid or magnesium oxide (MgO) addition is available in the program now on a force-through basis. The user supplies an additive rate and the program calculates the equipment requirements. Models which relate the additive concentrations to SO_2 removal are still being developed.

PROGRAM DESCRIPTION

Design Basis

The Shawnee lime-limestone scrubbing program is developed to project a complete conceptual design package for lime or limestone scrubbing utilizing a TCA, venturi - spray tower, or spray tower absorber with any of four sludge disposal options. Basic design parameter ranges are shown below.

Plant size	100-1300 MW
Coal sulfur	2%-5% (1500-4000 ppm SO_2)
Scrubber gas velocity	8-12.5 ft/sec
Liquor recirculation rate	25-120 gal/ACF (at scrubber outlet)
Slurry residence time	2-25 min
Scrubber slurry solids	5%-15%
Reheat (steam)	Maximum 225°F

Results for conditions outside these design ranges are not necessarily invalid, but are subject to potential reduced accuracies.

Process Description

Flow diagrams for the basic processes are shown in Figures 1 to 3. A general description of these processes follows.

Raw Material Handling and Preparation. For the limestone slurry process the raw material handling and preparation area includes equipment for receiving limestone by truck or rail, a storage stockpile, live in-process limestone storage equipment, and the equipment for crushing and wet grinding the raw limestone to the desired size for feed to the scrubbers. The product slurry from the mills is pumped to a slurry feed tank adjacent to the scrubbing area for distribution to the scrubbers.

For the lime slurry process the raw material handling and preparation area (see Figure 4) includes equipment for conveying lime from a calcining plant located 1500 feet from the area to a storage silo and slaking the lime. The product slurry from each of the slakers overflows to a slurry receiving tank from which it is pumped to a common slurry feed tank. The slurry is then pumped to the scrubbing area for distribution to the scrubbers.

Flue Gas Handling. Flue gas from the power unit ducts is fed to a common plenum, from which any number of scrubber trains can be fed. To minimize the problems associated with gas distribution for such a system, separate fans are included on each side of the plenum. The power plant fans are conventional induced-draft fans for balanced-draft boilers. The forced-draft scrubber fans are designed to overcome the pressure drop of the pollution control facilities. Particulate removal may optionally be included in the process upstream of the SO_2 absorbers. As an alternate, removal of the particulates within the scrubber system may also be evaluated.

Flue gas from the scrubber fans is contacted with a lime or limestone slurry in the absorber, which is equipped with a chevron-vane mist elimination system designed for upstream and downstream wash with fresh

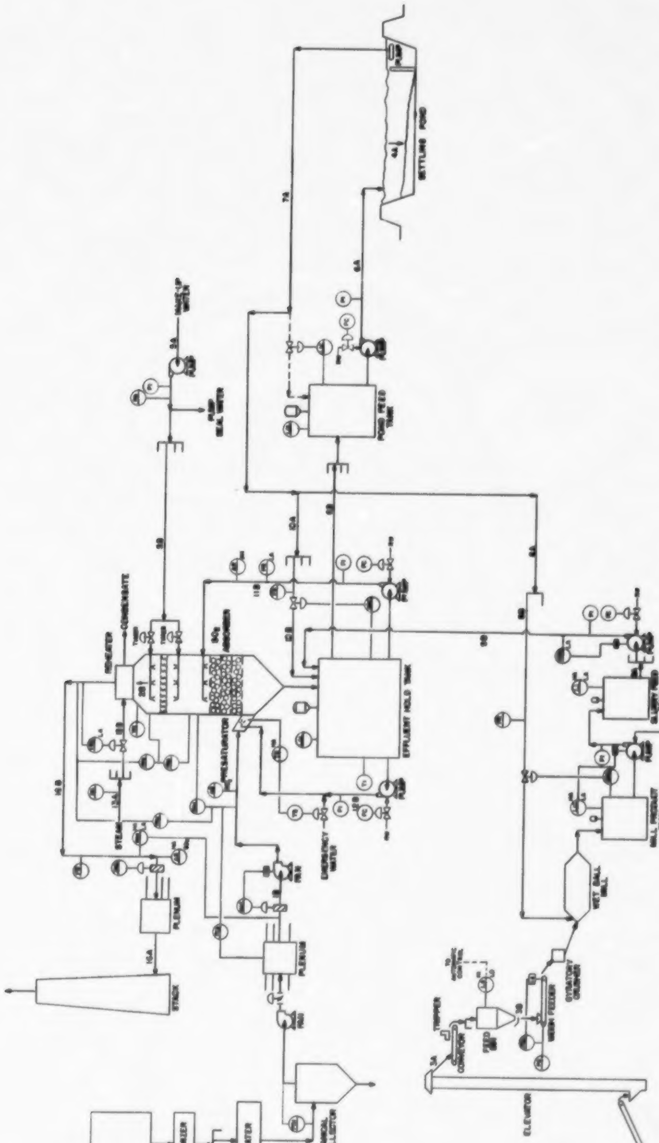

SETTLING POND
POND FEED TANK
MAKE-UP WATER
PUMP SEAL WATER
REHEATER
CONDENSATE
STEAM
SO_2 ABSORBER
PRESATURATOR
EFFLUENT HOLD TANK
EMERGENCY WATER
PLENUM
STACK
SLURRY FEED TANK
MILL PRODUCT TANK
WET BALL MILL
VIBRATORY UNLOADER
WEIGH FEEDER
FEED BIN
CONVEYOR
TRIPPER
ELEVATOR
MECHANICAL ASH COLLECTOR
AIR HEATER
ECONOMIZER
AIR
BOILER
PULVERIZED COAL
PILE
HOPPERS, FEEDERS & CONVEYORS
CRANE
SHAKER
PULLER

Figure 1. TCA - onsite ponding.

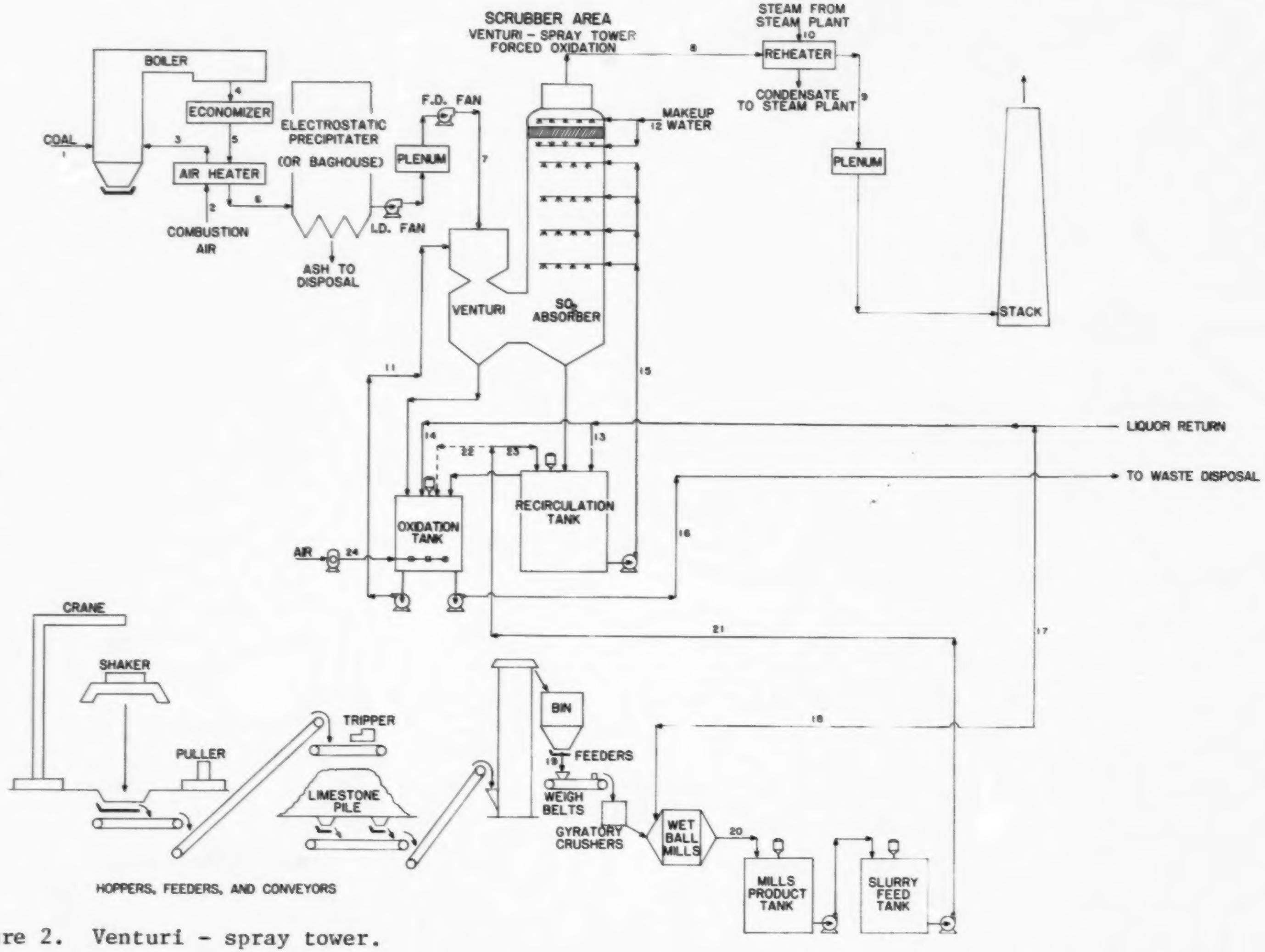

SCRUBBER AREA
VENTURI - SPRAY TOWER
FORCED OXIDATION
BOILER
COAL
1
ECONOMIZER
AIR HEATER
COMBUSTION AIR
2
3
4
5
6
ELECTROSTATIC PRECIPITATER
(OR BAGHOUSE)
ASH TO DISPOSAL
I.D. FAN
PLENUM
F.D. FAN
7
VENTURI
SO_2 ABSORBER
MAKEUP WATER
12
8
STEAM FROM STEAM PLANT
10
REHEATER
CONDENSATE TO STEAM PLANT
9
PLENUM
STACK
11
13
14
15
16
17
18
19
20
21
22
23
24
LIQUOR RETURN
TO WASTE DISPOSAL
OXIDATION TANK
RECIRCULATION TANK
AIR
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
HOPPERS, FEEDERS, AND CONVEYORS
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK

Figure 2. Venturi - spray tower.

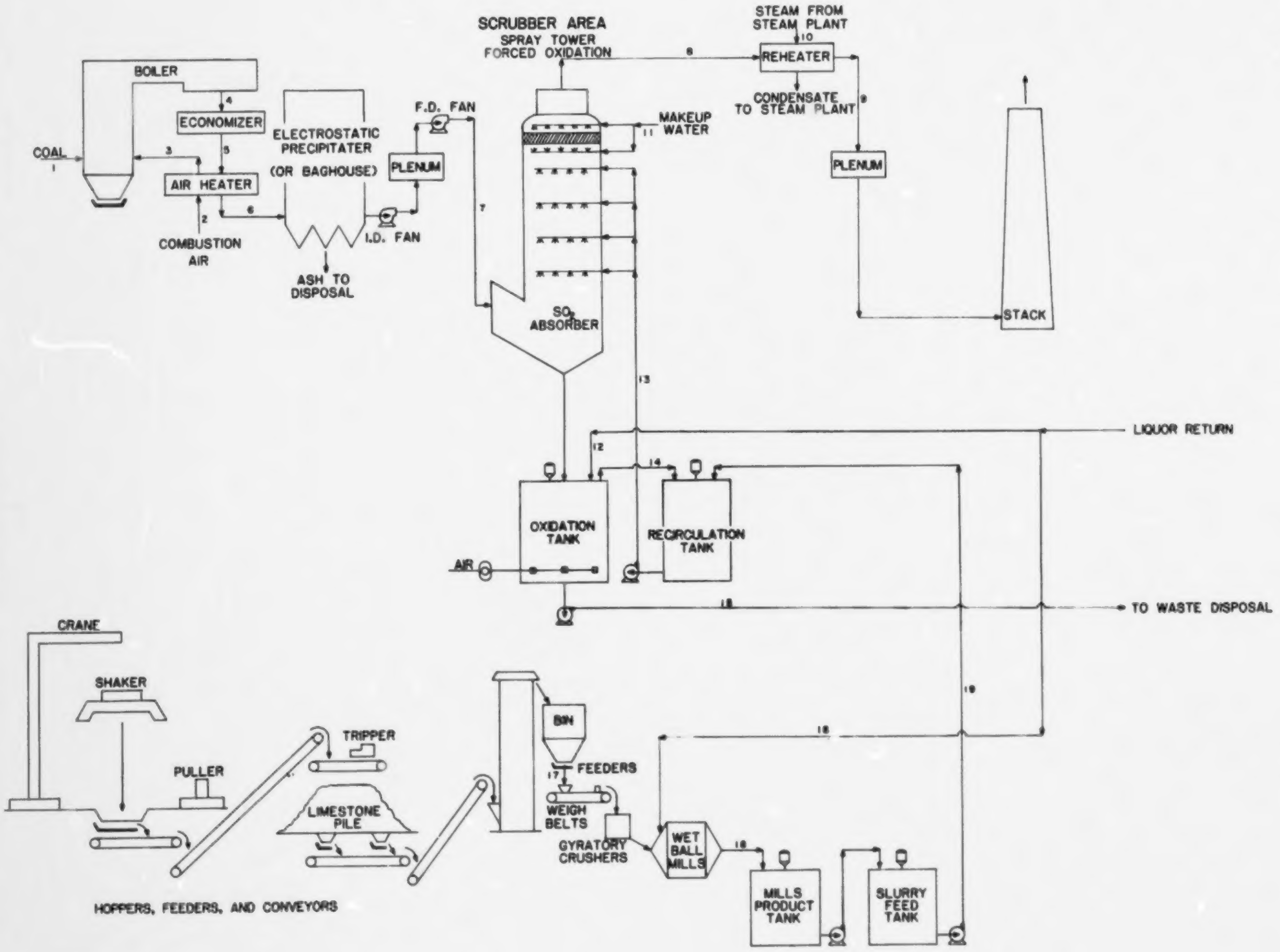

SCRUBBER AREA
SPRAY TOWER
FORCED OXIDATION
BOILER
ECONOMIZER
AIR HEATER
COAL
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
COMBUSTION AIR
ELECTROSTATIC PRECIPITATER (OR BAGHOUSE)
ASH TO DISPOSAL
I.D. FAN
F.D. FAN
PLENUM
MAKEUP WATER
SO_2 ABSORBER
STEAM FROM STEAM PLANT
REHEATER
CONDENSATE TO STEAM PLANT
PLENUM
STACK
LIQUOR RETURN
AIR
OXIDATION TANK
RECIRCULATION TANK
TO WASTE DISPOSAL
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
HOPPERS, FEEDERS, AND CONVEYORS
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK

Figure 3. Spray tower.

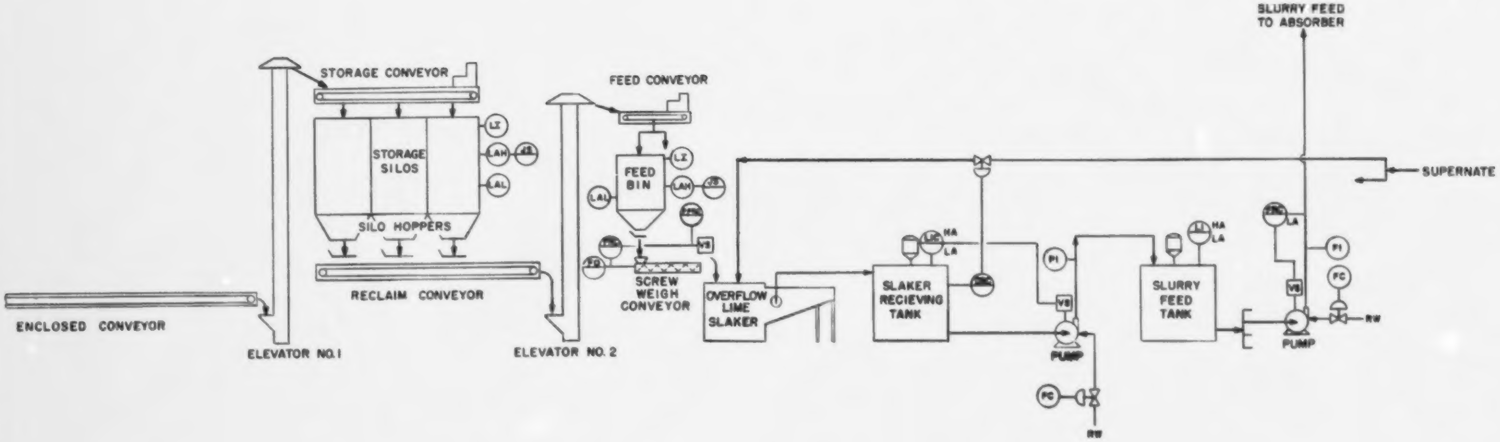

SLURRY FEED TO ABSORBER
STORAGE CONVEYOR
FEED CONVEYOR
STORAGE SILOS
LI
LAH
LAL
SILO HOPPERS
FEED BIN
SUPERNATE
RECLAIM CONVEYOR
SCREW WEIGH CONVEYOR
ENCLOSED CONVEYOR
ELEVATOR NO.1
ELEVATOR NO. 2
OVERFLOW LIME SLAKER
SLAKER RECIEVING TANK
HA
LA
PI
VS
PUMP
FC
RW
SLURRY FEED TANK

Figure 4. Lime slurry option absorbent preparation area.

makeup water. The outlet gas from the scrubber is reheated to the desired temperature by indirect steam reheat and discharged to the stack plenum.

Slurry Recirculation Loop. Makeup lime or limestone slurry from the slurry feed tank and recycled supernate or filtrate from the waste disposal area are fed to the absorber hold tanks where they are blended with the slurry draining from the absorber. The slurry is then recirculated to the absorber, venturi, or presaturator depending on the process selected. A bleed stream is pumped to the waste disposal area where the sludge is dewatered. The supernate or filtrate is returned to the scrubber area and raw material preparation area.

Waste Disposal System. The program has four alternate waste disposal options as shown in Figures 5 and 6. These options are listed below.

1. Onsite pond
   a. Unlined pond
   b. Clay-lined pond (cost and depth of clay lining is input)
   c. Synthetic-lined pond (cost of liner is input)
2. Thickener - pond
3. Thickener - fixation fee
4. Thickener - filter - fixation fee

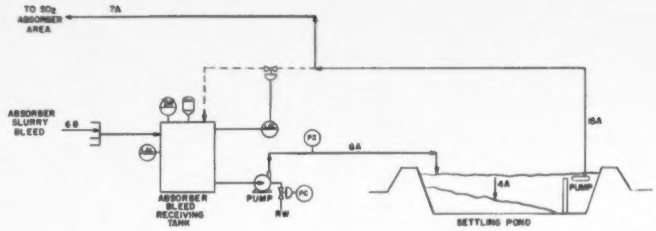
The onsite ponding options may also be run with fixation fees applied to them. For alternatives 3 and 4, the fixation fee must include costs for transportation and disposal of the fixed sludge offsite. For alternatives 1 and 2, however, only the costs for fixation need be provided since the fixed sludge can be disposed of at the existing pond site.

For the ponding alternatives the program allows for the onsite pond to be sized larger or smaller than the normal projected lifetime capacity. This option has been incorporated (1) to account for variations in the sulfur content of fuel, (2) to evaluate design philosophy in construction of ponds for less than the total amount of sludge to be disposed of (this requires assessment of additional cost for expanding pond later), or (3) to allow the feed preparation and scrubbing areas to be sized based on maximum sulfur contents expected, while sizing the pond based on average sulfur contents. An example output of the onsite ponding model is shown in Table 1.

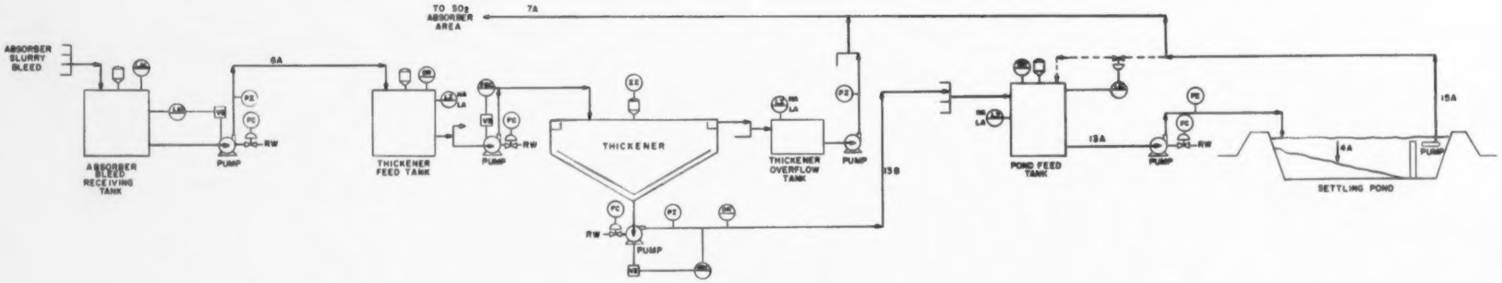
Program Models

The overall program provides specific output for the following items.

1. Detailed material balance
2. Particulate removal equipment and cost
3. Major equipment specifications and costs
4. Capital investment requirements
5. Annual and lifetime revenue requirements

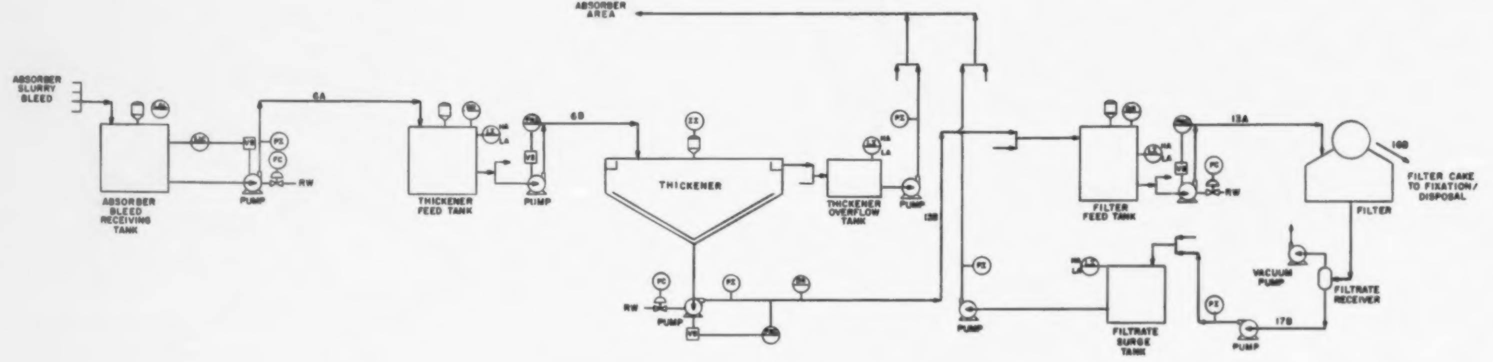

TO SO2 ABSORBER AREA
7A
ABSORBER SLURRY BLEED
6B
ABSORBER BLEED RECEIVING TANK
PUMP
RW
PI
PC
6A
15A
4A
PUMP
SETTLING POND

ONSITE PONDING

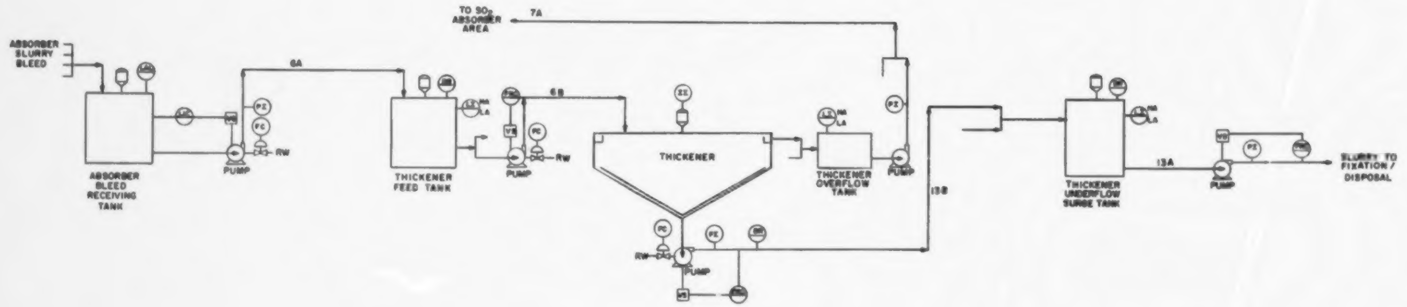

TO SO2 ABSORBER AREA
7A
ABSORBER SLURRY BLEED
ABSORBER BLEED RECEIVING TANK
PUMP
RW
6A
THICKENER FEED TANK
PUMP
RW
THICKENER
RW
PUMP
THICKENER OVERFLOW TANK
PUMP
13B
POND FEED TANK
13A
PUMP
RW
15A
4A
PUMP
SETTLING POND

THICKENER--PONDING

Figure 5. Onsite pond disposal alternatives.


ABSORBER SLURRY BLEED
ABSORBER BLEED RECEIVING TANK
PUMP
RW
6A
THICKENER FEED TANK
PUMP
6B
THICKENER
THICKENER OVERFLOW TANK
PUMP
13B
TO SO2 ABSORBER AREA
PUMP
FILTER FEED TANK
13A
FILTER
16B
FILTER CAKE TO FIXATION/ DISPOSAL
VACUUM PUMP
FILTRATE RECEIVER
FILTRATE SURGE TANK
PUMP
17B

SLUDGE DISPOSAL OPTION 4--THICKENER--FILTER--FIXATION(FEE)


ABSORBER SLURRY BLEED
ABSORBER BLEED RECEIVING TANK
PUMP
RW
6A
THICKENER FEED TANK
PUMP
6B
TO SO2 ABSORBER AREA
7A
THICKENER
PUMP
THICKENER OVERFLOW TANK
PUMP
13B
THICKENER UNDERFLOW SURGE TANK
13A
PUMP
SLURRY TO FIXATION / DISPOSAL

SLUDGE DISPOSAL OPTION 3--THICKENER--FIXATION(FEE)

Figure 6. Fixation alternatives.

Table 1. POND DESIGN EXAMPLE PRINTOUT

```
                    POND DESIGN
                    ---- ------

          OPTIMIZED TO MINIMIZE TOTAL COST PLUS OVERHEAD

(POND DESIGNED FOR  85.00 % OF PROJECTED LIFETIME CAPACITY)


          POND DIMENSIONS
          ---- ----------

          DEPTH OF POND                      21.43 FT
          DEPTH OF EXCAVATION                 3.16 FT
          LENGTH OF PERIMETER            14748.    FT
          LENGTH OF DIVIDER               2693.    FT

          AREA OF BOTTOM                  1355.    THOUSAND YD2
          AREA OF INSIDE WALLS             149.    THOUSAND YD2
          AREA OF OUTSIDE WALLS            113.    THOUSAND YD2
          AREA OF POND                    1494.    THOUSAND YD2
          AREA OF POND SITE               1665.    THOUSAND YD2
          AREA OF POND SITE                344.    ACRES

          VOLUME OF EXCAVATION            1524.    THOUSAND YD3
          VOLUME OF SLUDGE TO BE         10233.    THOUSAND YD3
          DISPOSED OVER LIFE OF PLANT     6343.    ACRE FT


          POND COSTS (THOUSANDS OF DOLLARS)
          ---- ----- ---------- -- --------

                                      LABOR    MATERIAL    TOTAL
                                      -----    --------    -----

          CLEARING LAND                514.                 514.
          EXCAVATION                  2844.                2844.
          DIKE CONSTRUCTION           1037.                1037.
          LINING( 12. IN. CLAY)       1253.                1253.
          SODDING DIKE WALLS            65.         53.     118.
          ROAD CONSTRUCTION              9.         19.      27.
          ------------------------------------------------------
          POND CONSTRUCTION           5722.         72.    5794.
          LAND COST                                        1204.
          ------------------------------------------------------
          POND SITE                                        6998.
          OVERHEAD                                         3615.
          ------------------------------------------------------
          TOTAL                                           10613.
```

General descriptions of the models for these items are given below.

Material Balance Model. This model performs the material balance calculations for the process flow streams in a limestone or lime system. The outputs of the material balance model consist of:

Hot gas to scrubber--

Total flow rate, composition, and temperature
Flow rate of each gaseous constituent
Hot gas humidity and wet-bulb temperature
Fly ash loading and flow rate
Corresponding coal firing rate

Wet gas from scrubber--

Total flow rate, composition, and temperature
Flow rate of each gaseous constituent
Wet gas saturation humidity
Fly ash loading and flow rate
Water pickup by flue gas (humidification and entrainment)
SO_2 concentration

Flue gas to stack--

Total flow rate, composition, and temperature
Flow rate of each gaseous constituent
Fly ash loading and flow rate
SO_2 removal
SO_2 emission

Steam reheater (in-line)--

Superficial gas velocity (face velocity)
Required heat input
Steam consumption
Gas pressure drop
Overall heat transfer coefficient
Reheater area requirement per train
Reheater pipe diameter and length
Number of pipes per bank and number of banks (square pitch) per train

Water balance--

Available water
Water required
Makeup water requirements

Scrubber system--

Total number of scrubbing trains
SO_2 removal efficiency
Particulate removal in scrubber
Particulate removal efficiency
Flue gas pressure drop across TCA beds, venturi, or spray tower
Total system pressure drop
Liquid to gas ratio (specified or calculated)
Limestone or lime addition rate
Limestone or lime stoichiometry
Fly ash CaO stoichiometry
MgO stoichiometry (from fly ash and separate addition, if any)
Adipic acid concentration
Scrubber inlet slurry pH
Makeup water requirement
Scrubber cross-sectional area per train

Solids disposal system--

Clarifier cross-sectional area

System sludge discharge--

Total flow rate and composition
Liquor pH and dissolved solids

Flow rates and compositions of system internal flow streams--

Scrubber slurry bleed
Total supernate return
Supernate to wet ball mill
Supernate to scrubber or effluent hold tank
Limestone or lime slurry feed
Recycle slurry to scrubber
Flue gas cooling slurry or recycle slurry to venturi
Clarifier underflow slurry
Clarifier overflow supernate
Filter cake slurry
Filtrate from filter

Particulate Removal Model. A model for sizing and predicting the flange-to-flange installed costs of several particulate removal devices is included in the program. Costs for a cold ESP, a hot ESP, a baghouse, and a wet scrubber are predicted based on the gas flow rate, particulate removal (%) desired, and coal sulfur levels. At present the model does not include costs for the ash disposal system or ductwork to and from the unit. The printout for this model is separate from the FGD capital investment so that the user may keep the SO_2 removal and the particulate removal charges separated. In the near future an option will be available in the program in which the particulate and SO_2 removal costs may be combined for both investment and operating requirements. An example of the particulate model may be seen in the example printout shown in Table 2.

Major Equipment Specification Model. Based on results from the material balance model and some user-supplied variables this model sizes and prices the major equipment and provides equipment specifications by area. The equations for predicting equipment costs were updated in 1979 to reflect current costs. These equations will periodically be updated based on future cost data.

The following procedures are utilized for determining the size or specifications of the major process equipment in the limestone slurry process raw material handling and preparation area.

Gyratory crushers--Two parallel 50% capacity gyratory-type crushers are utilized to reduce the inlet stone size from minus 1-1/2 inch to minus 3/4 inch for feed to the ball mills.

Ball mills--The grinding mills are rubber-lined, open-circuit, overflow wet ball mills that have a 30% ball charge and produce a 60% slurry. The number of ball mills is determined by total mill horsepower (calculated from the limestone throughput rate specified in the material balance) and the fineness of grind and limestone hardness factors (which are program inputs). The fineness of grind index factor is related to the desired particle size distribution of the ground limestone. One-mill systems are used for horsepower less than 200 and two parallel mill systems for horsepower between 200 and 5000. For horsepower greater than 5000, the number of parallel mill systems is determined assuming a maximum mill size of 2500 horsepower.

The lime slurry process raw material handling and preparation area utilizes the following major process equipment.

Lime conveyor--One enclosed belt conveyor is used to transport lime 1500 feet from an across-the-fence lime calcination plant to the storage silos. The conveyor operates at belt speeds of 100 to 300 ft/min and handles up to 75 tons/hr of lime.

Lime storage silo--A 15-day dead storage capacity is used to calculate the volume of the lime storage silos. The silos are concrete, with the height of the actual storage section of the silo assumed to be one and a half times the diameter. Total height of the silo is equal to the height of the actual storage section plus the height of the carbon steel hopper plus 5 feet. Parallel storage silos are used for storage volumes greater than the capacity of the largest silo (147,200 ft^3).

Lime slaker--Lime is slaked at slurry concentrations of 20% to 25% solids in dual-compartment, overflow slakers which can be designed with

Table 2. PARTICULATE MODEL EXAMPLE PRINTOUT

```
              INDUSTRIAL BRIEFING VST XX

                         PARTICULATE REMOVAL INVESTMENT AND OPERATING COST
   WPSUL CONTENT (%):            3.12          PARTICULATE EMISSICN REGULATION (LB ASH/MILLION BTU):   0.06
   ASH CONTENT (%):             16.00          FLUE GAS TEMPERATURE (COLD) (F):                       300.0
   BTU RATING:                  10500          FLUE GAS TEMPERATURE (HOT) (F):                        700.0
   BOILER TYPE:                 DRY PULVERIZED COAL  COST OF ELECTRICITY ($/KWHR):                     0.03
   NO. OF SCRUBBERS:                4          COST OF STEAM ($/THOUSAND LB):                          2.00
   SCRUBBER VELOCITY (FT/M):    600.0          FIRST YEAR CAPITAL CHARGE FACTOR:                      0.194
   PLANT SIZE (MW):               500          BAGHOUSE RATIO (OPER. SQ.FT./ACTUAL SQ.FT.):            0.80
   OPERATING HRS/YR:             4512          BAG COST ($/SQ.FT.):                                    0.70
   PUMPING RATE (GAL/1000 ACF):  20.00         BAG LIFE(YEARS):                                        3.00
   SCA RATIO:                    1.100         FLUE GAS REHEAT TEMPERATURE (F):                        175.
     (ACTUAL SQ.FT./CALC. SQ.FT.)              CHEMICAL ENGINEERING PLANT INDEX:                      265.0


                                   ELECTROSTATIC PRECIPITATORS

                                        COLD           HOT         BAGHOUSE FABRIC FILTERS     SCRUBBERS
 REQUIRED REMOVAL EFFICIENCY (%):       99.50         99.50
 DRIFT VELOCITY (FT/M):                 25.78         20.00              99.50                  99.50
 SPECIFIC COLLECTION AREA (SQ.FT./ACFM): 226.05       291.41
 COLLECTION AREA (SQ.FT.):           348077.3      684896.3
 TOTAL CORONA POWER (KW):               467.4         713.4            721033.4
 AUXILIARY POWER (KW):                  293.0         620.5
 FAN POWER (KW):                        241.0         367.8              461.5
 PUMP POWER (KW):                                                       1204.9                 8998.3
 TOTAL POWER (KW):                     1001.4        1701.8                                    3796.0
 OPERATING AIR/CLOTH RATIO:                                             1666.3                12794.3
 INSTALLED AIR/CLOTH RATIO:                                                2.7
 REQUIRED PRESSURE DROP (INCHES):         1.0           1.0                2.1
 DIAMETER (FEET):                                                          5.0                   37.3
 REQUIRED REHEAT (BTU/HR):                                                                         32
 STEAM SUPPLY/YR (THOUSAND LB):                                                            58660416.0
 INSTALLED COST (1979 DOLLARS):     $ 5046359     $ 8923927                                  294084.2
 FIRST YEAR CAPITALIZED COST:       $  979162     $ 1731539          $ 13998391            $ 21061472
 ANNUAL POWER COST:                 $  131027     $  222672          $  2716154            $  4086627
 ANNUAL OPERATING AND                                                $   218038            $  1674110
  MAINTENANCE COST (1979 DOLLARS):  $   95696     $  139421
 REPLACEMENT COST (1979 DOLLARS):                                    $    70773            $  1105205
 ANNUAL REHEAT COST:                                                 $   220712
                                                                                           $   588168


 TOTAL ANNUAL COST:                 $ 1205885     $ 2093632
 ANNUALIZED COST OF POWER(MILLS/KWHR):   0.53          0.93          $  3225677            $  7454110
                                                                          1.43                   3.30
```

slaking capacities of up to 33 tons/day. Parallel slaking trains are used for larger lime capacities. The number and size of parallel slakers required are determined based on the capacity of the largest size slaker available (33 tons/day).

The following procedures are utilized for determining the size or specifications of the major process equipment in the scrubbing area.

Scrubbing trains--The number of parallel scrubbing trains is either an input to the program or is established as an override to the input value based on the minimum number of scrubber trains required. The minimum number of trains is calculated considering the saturated flue gas velocity and volumetric flow rate at the scrubber outlet in conjunction with the maximum cross-sectional area assumed for the TCA scrubber (1008 ft^2). Flue gas and slurry recirculation rates per train are calculated by dividing the total flow rates from the overall material balance model by the number of operating scrubbing trains.

Scrubbers--Scrubber cross-sectional area is calculated considering the outlet flue gas rate per train in conjunction with the specified scrubber design gas velocity. Number of scrubber grids, beds, and height of spheres per bed are inputs to the program. The height of the scrubber is assumed to remain constant for all scrubber sizes and internal configurations. A presaturator compartment is included at the scrubber inlet for the TCA and spray tower and chevron-type mist eliminators near the outlet. Materials of construction for the scrubbers and internals are listed below.

Venturi:	Carbon steel acid-resistant lining
Shell:	Rubber-lined carbon steel
Grids:	Type 316L stainless steel
Spheres:	1-1/2-inch-diameter, Nitrile foam

Mist eliminator, slurry header, and nozzles: Type 316L stainless steel

Reheaters--Reheater cross-sectional area is calculated based on the superficial gas velocity which is input to the program and the volumetric gas flow rate per train at scrubber outlet conditions. Reheater surface area requirements are calculated in two steps: (1) surface area requirements for reheat to 150°F and (2) requirements for reheat to the specified reheat temperature. The portion of the reheater tubes required to reheat to 150°F are Inconel and the remaining tubes are Cor-Ten. Reheater design and costs are based on use of 1-inch tubes on a 2-inch square pitch.

Fans--The fans are centrifugal (double width, double inlet) with radial impellers and are constructed of carbon steel. They are equipped with variable-speed fluid drives. Fan horsepower is calculated based on the inlet gas flow rate per train and the calculated pressure drop for the scrubber, mist eliminator, reheater, and duct.

The following procedures are utilized for determining the size or specifications of the major process equipment for the various waste disposal area alternatives.

Thickeners--The thickeners are constructed of flake-glass-lined carbon steel walls with 1-foot-thick concrete conical basins and are equipped with rake mechanisms. A concrete underflow tunnel including pumps and piping for transferring the slurry is included. Total thickener cross-sectional area is calculated by the material balance portion of the model as a function of the settling rate and settled solids density, which are inputs into the program, and the quantity of sludge in the effluent slurry calculated in the material balance model. The number of thickeners required is determined assuming a maximum thickener diameter of 400 feet. Thickener height is calculated as a function of the diameter.

Filters--Rotary drum filters constructed of carbon steel and equipped with a vacuum pump, a filtrate pump, and a vacuum receiver are utilized. Filter size is presently determined as a function of the filtration rate, which is a program input, expressed in tons of dry solids/ft^2/day in conjunction with the total quantity of sludge. The minimum and maximum sizes of filters considered have effective filtration areas of 50 and 900 ft^2 respectively. Single filters are used up to required filtration areas of 100 ft^2. For total filtration areas between 100 and 1800 ft^2, two parallel filters are assumed. For total filtration areas greater than 1800 ft^2, the number and size of parallel filters required are determined based on the capacity of the largest filter size.

The size or specifications of tanks, agitators, and pumps for each of the areas are determined by utilizing the following procedures.

Tanks--Tank volume is calculated based on the residence time, which is either a program input or assumed. An additional 10% volume is added for freeboard. All tanks are constructed of carbon steel and the slurry tanks are flake glass lined. Except for the absorber bleed receiving tanks and the thickener overflow tanks, each tank is designed with diameter equal to height up to a maximum height of 60 feet. For tanks larger than 60-foot diameter, tank height is fixed at 60 feet and diameter is calculated. Absorber bleed receiving tank height is equal to the effluent hold tank height and the diameter is calculated. Thickener overflow tank height is set equal to the height of the thickener and the diameter calculated. As an override to the calculated diameter a minimum diameter equal to one-half the height is fixed for all tanks. The thickener and filter feed tanks are not used unless more than one thickener or filter is required.

Agitators--All slurry tanks are equipped with a four-blade, pitched-blade, turbine agitator. Agitator horsepower requirements are calculated on the basis of total torque, which is a function of the degree of agitation required (expressed as torque/unit volume), total tank volume, tank diameter, and the slurry specific gravity. Unit torque (torque/unit volume) for each tank is determined as a function of the percent solids in the slurry.

Slurry pumps--All slurry pumps are rubber lined, centrifugal with water seals, and are equipped with either a variable- or constant-speed drive. Pumps are usually spared, with the number of operating pumps determined by the maximum available pump size of 20,000 gpm.

Water pumps--Vertical, multiple-stage, turbine makeup-water pumps capable of providing a static head of 200 feet are provided for each 10,000 gpm of water required. The pumps are carbon steel and spared.

Capital Investment Requirement Model. A detailed list identifying the required inputs to the program is included in the users manual. To facilitate projection of overall capital investment requirements and display of the results, the lime-limestone scrubbing process is divided into three distinct areas: (1) the raw material handling and preparation area, (2) scrubbing area, and (3) waste disposal area. The raw material handling and preparation area includes costs for all of the lime-limestone receiving and preparation equipment and field materials up to and including the slurry feed tank. The scrubbing area includes costs for all of the flue gas equipment and field materials downstream of the air heater, up to but excluding the stack plenum, and all slurry equipment between the slurry feed tank and the pond feed tank. Costs for the mechanical collector may be included optionally. Costs for the power plant fan and the downstream plenum and stack are considered to be an integral part of the power plant and are therefore not included in the estimates. The waste disposal area includes costs for the equipment and field materials downstream of the pond feed tank including those associated with the supernate return pumps and piping.

Costs for field materials are based on the materials of construction or specifications discussed below.

Piping--Carbon steel pipe and gate valves are used for all water lines including pond supernate. For slurry lines less than 3-inch diameter, stainless steel pipe is used, whereas for all larger size lines rubber-lined carbon steel piping is used. Stainless steel strainers are used for pipes less than 4-inch diameter and rubber-lined strainers are used for 4-inch-diameter and larger pipes.

For slurry lines less than 3-inch diameter, stainless steel plug valves are used. Eccentric plug valves are used for slurry lines between 3- and 20-inch diameter, and knife gate valves are used for lines greater than 20-inch diameter. Handwheel operators are used for valves less than 12-inch diameter and air cylinder actuators for larger valves. Typical piping layouts are assumed as functions of flow capacities and the number of trains and costs are correlated to flow rates in gal/min. Control valve costs are included in instrumentation. Costs are included for a rubber-lined downcomer from the TCA scrubber to the effluent hold tank and a spare slurry disposal line to the pond.

Ductwork--Costs are included for the inlet plenum and all ductwork between the inlet and stack plenums including insulation. Costs for the stack plenum are not included since it is required regardless of the method for removing SO_2 from the gas. Stack plenum elevation is set equal to effluent hold tank height with a minimum elevation of 20 feet for small hold tanks. Each scrubber train includes two dampers with costs included for expansion joints.

Materials of construction for all ductwork is Cor-Ten with the exception of ductwork between the scrubber and reheater outlet which is type 316 stainless steel. Duct size is based on a square cross-section and a nominal design velocity of 3000 ft/min at inlet conditions (approximately 300°F). Duct cross-sectional area downstream of the scrubbers and reheaters is set equal to the upstream cross-sectional area.

Foundations--Concrete foundations for each equipment item are fixed according to equipment sizes. Foundations for the structure are estimated on the basis of the weight of the structure.

Structural--Structural estimates are based on the structure arrangement shown in Figure 7. The total quantity of structure required for each scrubber train and the corresponding costs are related to effluent hold tank volume, scrubber cross-sectional area, and number of scrubbing trains.

Electrical--The electrical estimate is divided into four sections: (1) costs of feeder cables from the power plant transformer yard to field modules for each area, (2) transformer costs for each area, (3) costs of power supply from area field modules to individual motors, and (4) motor control costs between remote control center, field module location, and individual motors for each area. For each area, total connected motor horsepower is calculated for use in establishing costs for (1) and (2). Costs for (3) and (4) are based on individual motor sizes and number of connected motors. A typical layout is assumed for each area in reference to the power plant transformer yard, remote control center, and other areas.

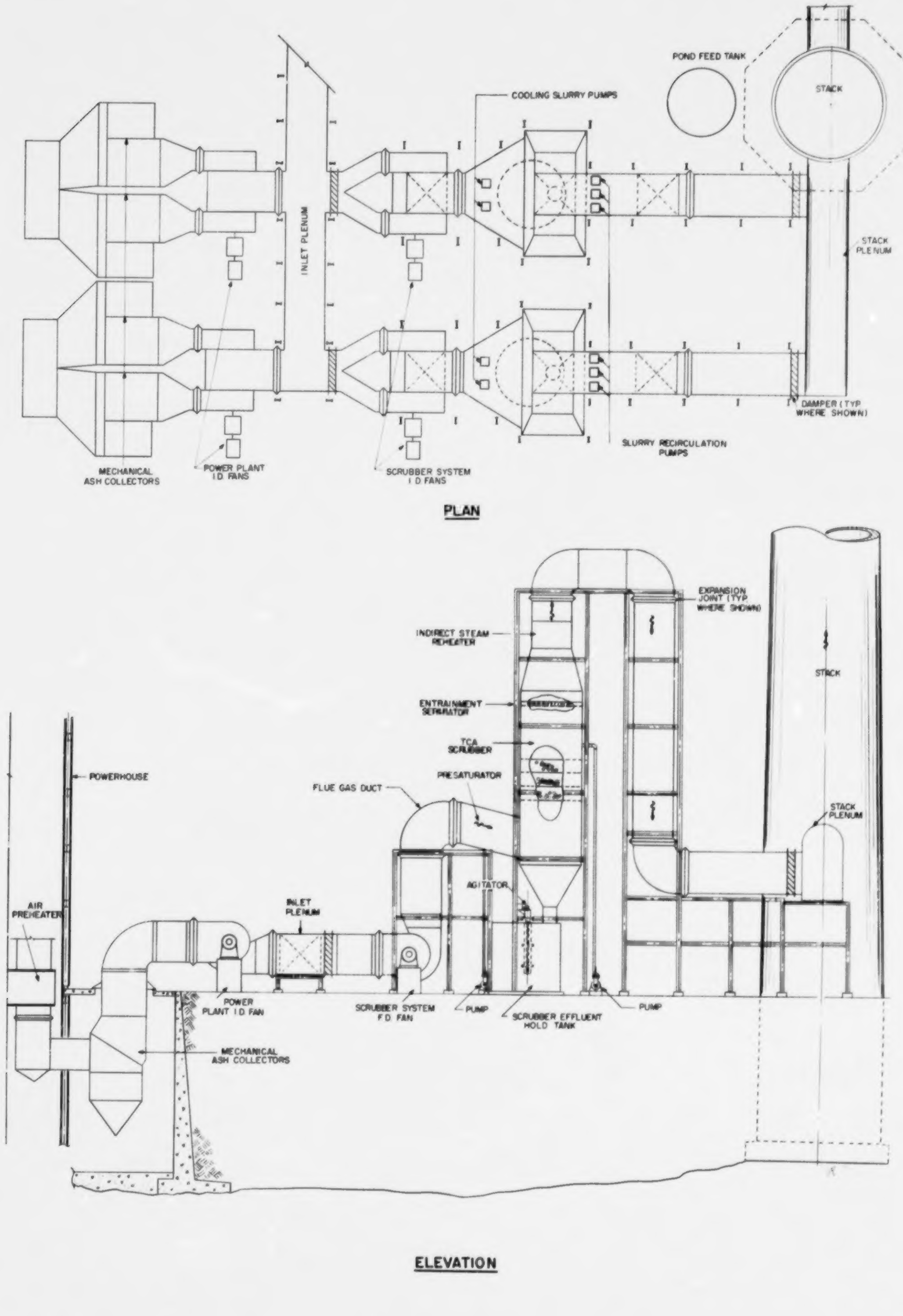

POND FEED TANK
STACK
COOLING SLURRY PUMPS
INLET PLENUM
STACK PLENUM
DAMPER (TYP WHERE SHOWN)
SLURRY RECIRCULATION PUMPS
MECHANICAL ASH COLLECTORS
POWER PLANT I.D. FANS
SCRUBBER SYSTEM I.D. FANS
PLAN
EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
STACK
ENTRAINMENT SEPARATOR
TCA SCRUBBER
PRESATURATOR
FLUE GAS DUCT
POWERHOUSE
STACK PLENUM
AGITATOR
INLET PLENUM
AIR PREHEATER
POWER PLANT I.D. FAN
SCRUBBER SYSTEM F.D. FAN
PUMP
SCRUBBER EFFLUENT HOLD TANK
PUMP
MECHANICAL ASH COLLECTORS
ELEVATION

Figure 7. TCA scrubber arrangement diagram.

Instrumentation--Instrumentation costs are based on (1) fixed costs for instruments which do not change in size and cost with equipment and pipe size variations and (2) variable costs for instruments which increase in size and cost as equipment and pipe sizes increase. Each of these costs may be dependent upon number of scrubbing trains, number of ball mills, number of pumps, etc. Costs are included for control valves, graphic boards and panelboards, annunciator, air dryers and piping, and instrument cable and wiring systems.

Buildings--The control room and motor control center are integrated with the power plant and prorated costs are included. For the limestone process, costs are included for one 90-foot-long by 30-foot-wide by 46-foot-high building to house the grinding facilities for a one-mill system. The width is increased by 30 feet for each additional mill.

Services and miscellaneous--Services and miscellaneous costs are based on the definition and costs given in Detailed Cost Estimates for Advanced Effluent Desulfurization Processes (EPA-600/2-75-006 or PB-242 541/1WP, January 1975). They are scaled proportionally to the relative power unit size ratio raised to the 0.6 power for all cases.

Pond construction--Disposal pond size is calculated based on a square configuration with a diverter dike three-fourths the length of one side. A pond construction diagram is shown in Figure 8.

A separate model is included to design and cost the onsite pond. The following design information is displayed:

Cubic yards of sludge to be disposed of over life of plant
Equivalent disposal volume in acre-feet
Pond depth in feet
Cubic yards of excavation required in construction of pond
Pond size in acres
Breakdown of surface area for the dike and pond

The pond model is based on either unlined, clay-lined, or synthetic-lined design and includes the following options in running the program:

Fixed-depth pond
Optimum-depth pond based on minimum pond investment
Optimum-depth pond based on minimum pond investment with available acreage and maximum excavation depth as overriding constraints

In addition to specifying pond design, the model also itemizes the breakdown of projected pond costs.

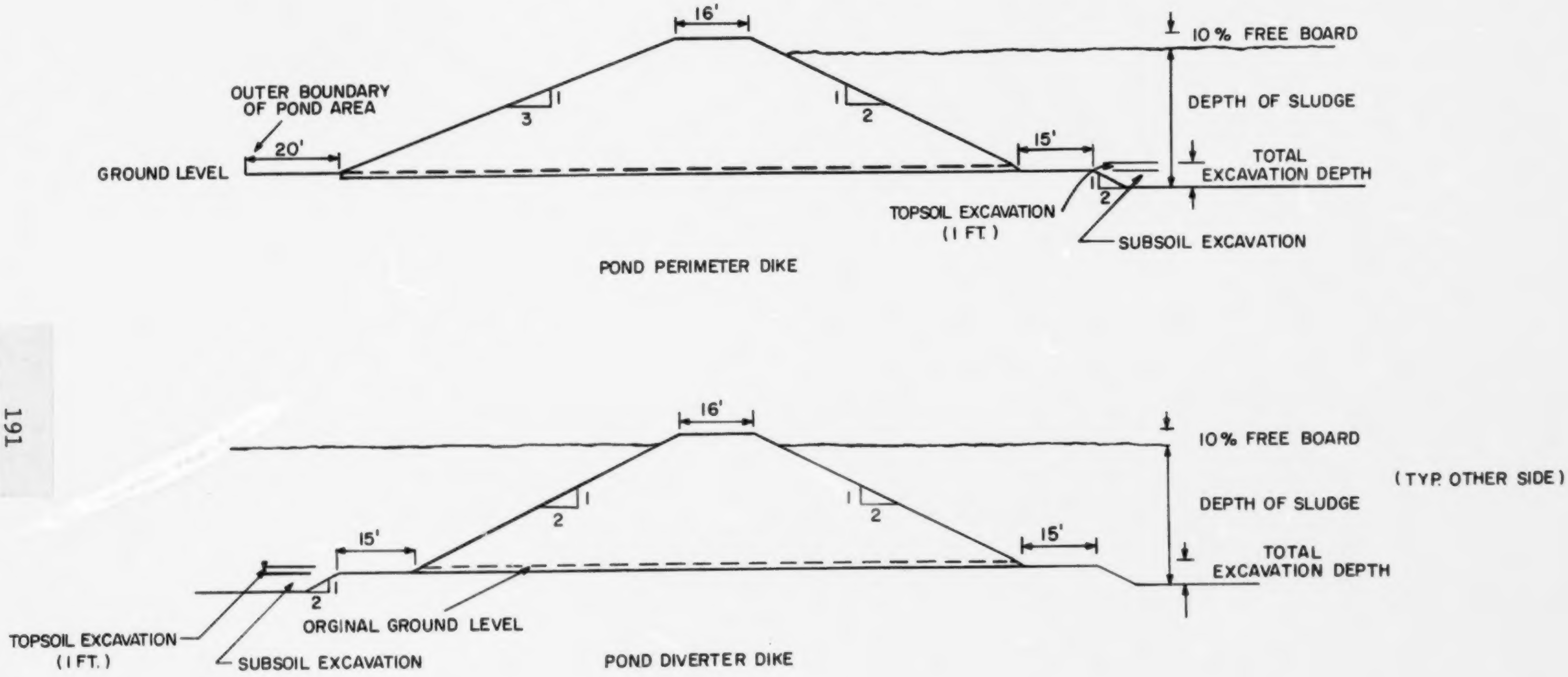

16'
10% FREE BOARD
OUTER BOUNDARY OF POND AREA
1
3
1
2
DEPTH OF SLUDGE
20'
15'
TOTAL EXCAVATION DEPTH
GROUND LEVEL
1
2
TOPSOIL EXCAVATION (1 FT.)
SUBSOIL EXCAVATION
POND PERIMETER DIKE
16'
10% FREE BOARD
(TYP. OTHER SIDE)
1
2
1
2
DEPTH OF SLUDGE
15'
15'
TOTAL EXCAVATION DEPTH
1
2
TOPSOIL EXCAVATION (1 FT.)
ORGINAL GROUND LEVEL
SUBSOIL EXCAVATION
POND DIVERTER DIKE

Figure 8. Pond construction diagram.

Annual and Lifetime Revenue Requirements Models. The TVA revenue requirement cost model incorporated into the overall program was initially developed to allow comparison of the revenue requirements of both recovery (product producing) and throwaway (nonproduct producing) processes. The output of this program has been modified only slightly for incorporation into the overall computer program. Descriptions of the procedures utilized within the model for calculating (1) the annual revenue requirements cost breakdown for the first year of operation of the plant and (2) the lifetime revenue requirements follow. The annual revenue requirements cost breakdown for the first year of operation of the plant is displayed in the same manner as average annual operating costs in Detailed Cost Estimates for Advanced Effluent Desulfurization Processes.

Annual revenue requirements direct costs--Raw material, process steam, and water requirements are transferred from the material balance model to the annual revenue requirements model. Operating motor horsepower requirements are transferred to the revenue requirements model from the capital investment model where the equivalent power requirement in kilowatts is calculated. In conjunction with the operating profile of the plant, which is presently fixed within the program, the annual quantities of raw materials and utilities are estimated. Operating labor and supervision are calculated by relating manning requirements to plant size (see Figure 9). Analyses are scaled from the quantities given in Detailed Cost Estimates for Advanced Effluent Desulfurization Processes. Maintenance rates are inputs to the program, with separate rates applied to (1) direct costs for the pond and (2) the total direct costs excluding the pond. These values, along with total capital investment requirements, are transferred to the revenue requirements model from the capital investment model. This procedure for considering maintenance allows the user flexibility to vary pond maintenance as functions of pond design (i.e., unlined, clay lined, or synthetic lined) while maintaining a constant maintenance rate for the remaining portion of the capital investment. In addition to displaying the annual quantities, this model displays the input unit costs for raw materials and utilities, the total annual cost for each raw material, and conversion cost items included in the direct costs.

Annual revenue requirements indirect costs--A detailed discussion of the indirect cost breakdown is included in Detailed Cost Estimates for Advanced Effluent Desulfurization Processes. Within the present model, straight-line depreciation is utilized and a depreciation period of 30 years is assumed. Depreciation is applied to the subtotal capital investment, thereby excluding the nondepreciable items (land and working capital). The cost of capital and taxes is based on undepreciated investment utilizing a capital charge percentage factor that is input to the program. The full value of capital charges for the first year of operation of the plant is reflected in the annual revenue requirements cost breakdown, which is displayed. Fixed rates for insurance (0.50% of total capital investment) and

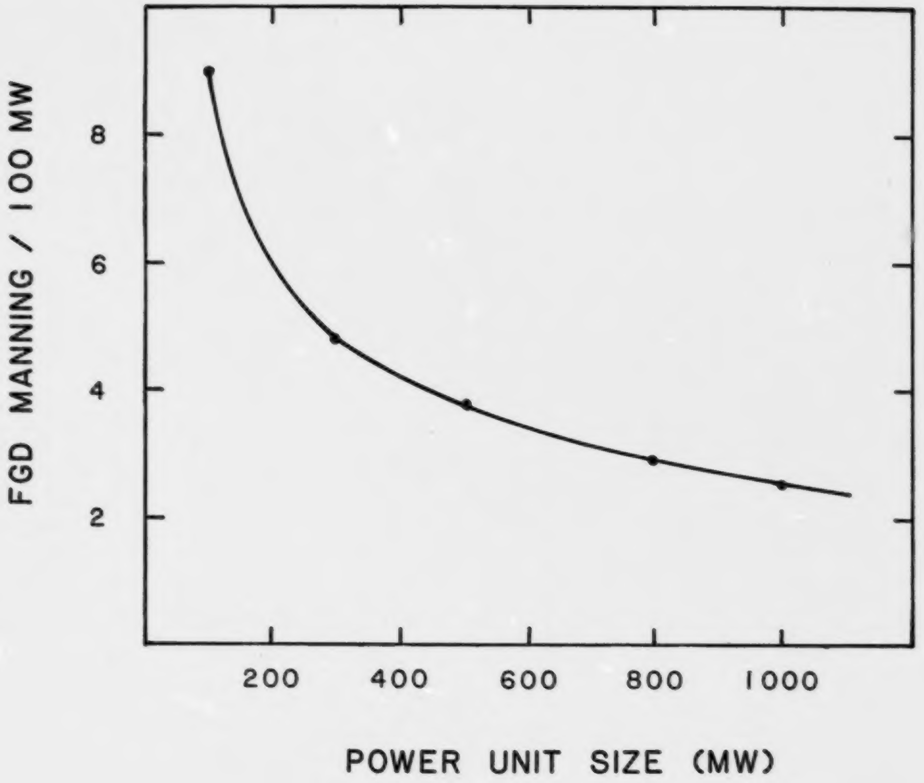


Figure 9. FGD manning.

interim replacements (0.67% of total capital investment) are incorporated in the program. The rates remain constant for each year of plant operation.

Plant and administrative overhead percentage factors are required inputs to the program. The input plant overhead factor is applied to conversion costs less utilities to estimate plant overheads, and the administrative overhead factor is applied to the operating labor and supervision costs to estimate administrative overheads.

Annual revenue requirements printout--The total annual revenue requirements are the sum of the subtotal direct and indirect costs. In addition to displaying the breakdown of the total annual revenue requirements, the input heat rate of the power unit in Btu/kWh, heat value of coal in Btu/lb, and the coal rate in tons/yr are displayed on the annual revenue requirements display sheet.

Lifetime revenue requirements--Projections of lifetime revenue requirements take into consideration both variations in capital charges and operating profile with plant age. The adjusted gross annual revenue requirements are calculated by the method illustrated above for the first year of operation, but for each year thereafter the model takes into consideration variations in capital charges and operating profile with plant age. The method of determining these factors is discussed below.

Capital charges--As discussed above, the portion of indirect costs reflecting the cost of capital and taxes is based on undepreciated investment. Since the basis for depreciation incorporated into the program is straight line, the cost of capital and taxes decreases uniformly over the life of the plant. The decrease in annual revenue requirements as a result of capital charge is therefore the same between years 29 and 30 as between years 1 and 2.

Operating profile--When the operating profile of the plant changes, the projected direct costs are modified to include the effect of increased or decreased operating capacity on costs. The current version of the program allows for the specification of three alternative operating profiles (see Figures 10 and 11), as indicated below, for projecting lifetime revenue requirements:

1. Profile similar to that utilized in the report Detailed Cost Estimates for Advanced Effluent Desulfurization Processes
2. Historical power plant operating profile based on FPC Form 67 data
3. Variable profile with annual load factors as input

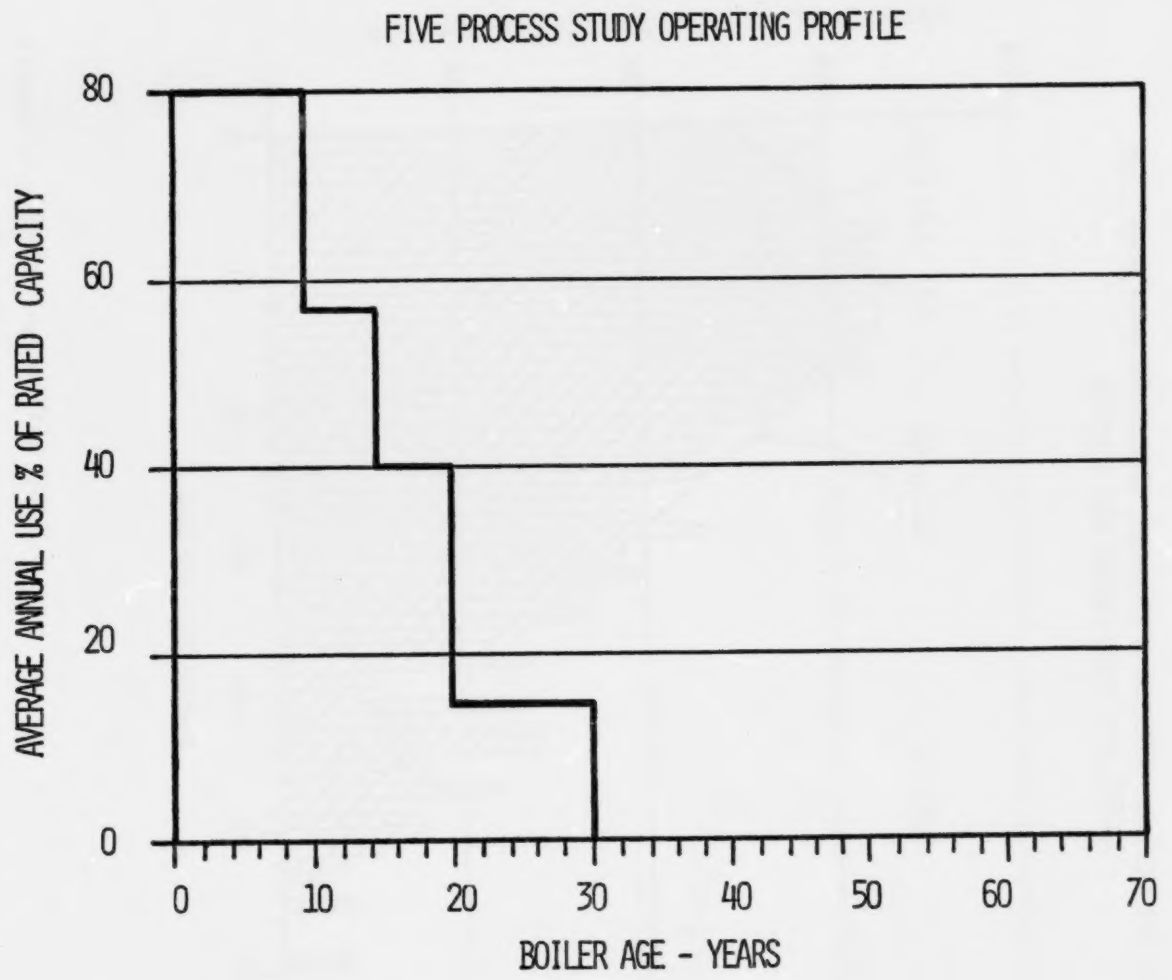


Figure 10. Operating profile No. 1.

FPC OPERATING PROFILE

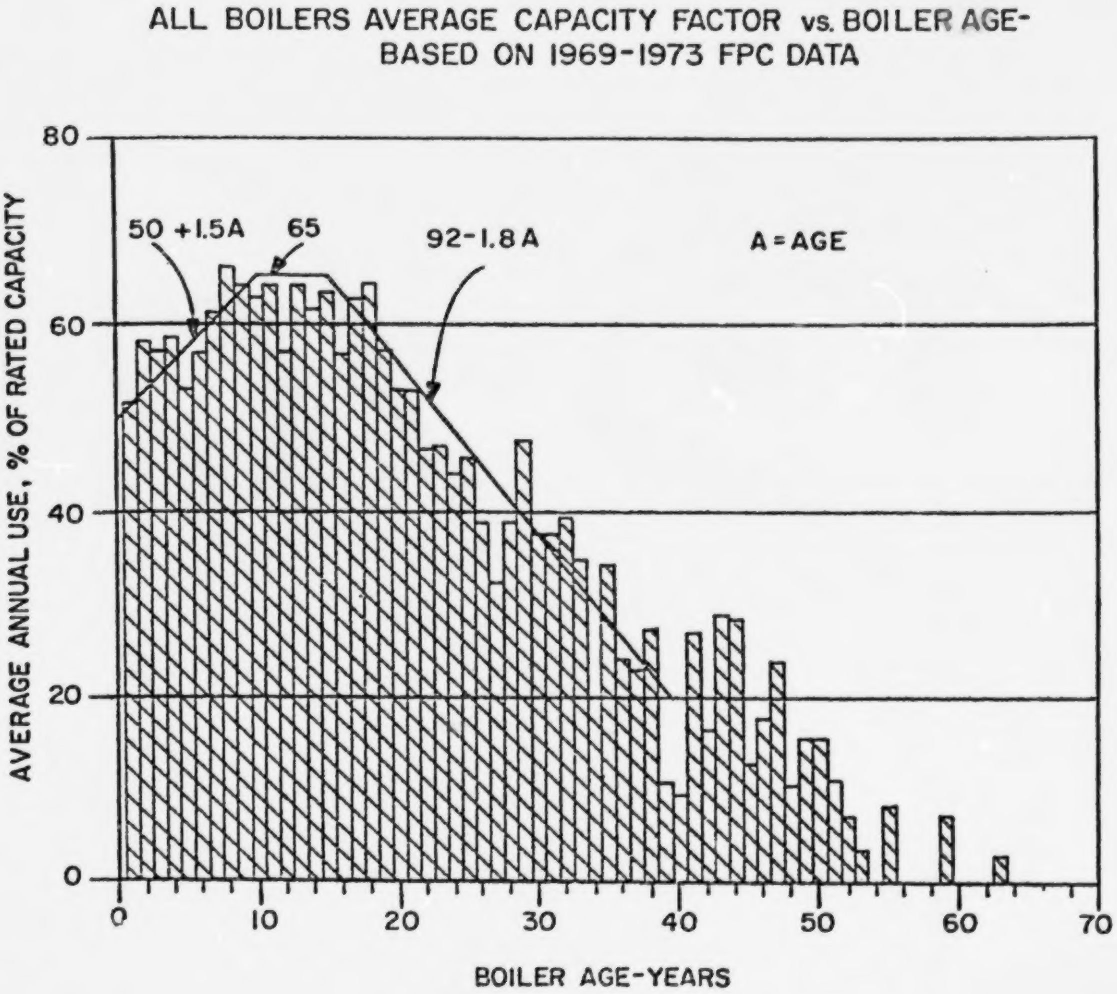


Figure 11. Operating profile No. 2.

Lifetime revenue requirement printout--For each year of operation of the plant, the lifetime revenue requirements model displays the load factor, heat requirement, fuel consumption, quantity of sulfur removed, equivalent byproduct rate, byproduct sales revenue in dollars per ton (which may be required in future), adjusted gross revenue requirements including regulated return on investment, sales revenue from salable byproducts, net increase in total revenue requirements, and cumulative net increase in total lifetime revenue requirements. Lifetime totals are shown for each of these items. Based on the totals displayed, the lifetime average unit revenue requirements for the SO_2 control facilities are calculated and displayed as:

Dollars per ton of coal burned,
Mills per kilowatthour,
Cents per million Btu heat input, and
Dollars per ton of sulfur removed.

The model also calculates the present worth of the equivalent revenue requirements in dollars by discounting annual revenue requirements at the cost of money to the initial year. Corresponding to the discounted revenue requirements over the life of the plant, the model calculates and displays the levelized increase in unit revenue requirements equivalent to the discounted revenue requirements and displays these unit costs on the same basis as described above. Levelized unit revenue requirements are defined by the following equation (P. W. = present worth).

$$\Sigma \text{ P. W. (annual revenue requirements)} =$$

$$\Sigma \text{ P. W. [(levelized unit cost)(number of units)]}$$

By defining levelized unit cost as a constant throughout the life of the plant, the equation is rearranged as follows:

$$\text{levelized unit cost} = \frac{\text{P. W. (annual revenue requirement)}}{\text{P. W. (number of units)}}$$

Therefore, the model for estimating lifetime revenue requirements also includes procedures for determining the present worth (discounting) of each unit (tons coal, kWh, etc.) for which levelized unit costs are displayed. These discounted units are divided into the discounted costs to obtain levelized unit costs.

FUTURE PROGRAM DEVELOPMENT

Further additions to the program are expected to be made as additional test data from Shawnee become available. At present the forced oxidation and additive models are of a force-through nature. The models for these options should be ready by early spring in 1980. Models for ductwork and ash disposal equipment for the particualte removal model should be ready in 3 to 4 months. Other options that may be added to the program in the future are (1) ID fan, (2) series scrubbers, (3) landfill costs, and (4) future equipment cost model updates.

PROGRAM USAGE

In its current state of development the program can be of use to utility companies or architectural and engineering firms involved in the selection and design of SO_2 removal facilities. The program also has potential for use by environmentalist groups or regulatory agencies. Although it is not intended to be used for projecting a final design, it can be used to assist in the evaluation of system alternatives prior to a detailed design. It should also be useful for evaluating the potential impact of various process variables on economics as a guide for planning.

Although the program was not meant to be used for comparing projected lime-limestone economics with economics for alternate processes, these comparisons should be valid as long as the basis for the alternate process economics is comparable to those included in the computer program for lime and limestone systems.

DEFINITIVE ECONOMICS OF ADVANCED LIMESTONE SCRUBBING PROCESSES

At present a definitive technical and economic assessment of the advanced limestone and lime scrubbing systems that have been tested at the EPA-sponsored Shawnee test facility is underway. The advanced systems include a TCA, a venturi - spray tower, and a spray tower arrangement. The evaluation will include a technical economic assessment of lime and limestone operation with and without forced oxidation, and with and without the chemical additives, adipic acid or magnesium oxide. The Shawnee Computer Economics program will be used extensively in preparing the sensitivity graphs for the report.

METHODS FOR USE OF THE PROGRAM

TVA has loaded a version of the Shawnee Lime-Limestone Computer Economics program onto the Control Data Corporation (CDC) national time-sharing computer system. An outside user may access the computer via an interactive time-sharing terminal after the appropriate billing arrangements have been made with CDC. Users who do not have computer capabilities may contact TVA and make arrangements for obtaining specific computer runs directly from TVA. One other option a potential user has for using the program is to request a tape copy of the program which TVA will provide along with available documentation for running it.

CURRENT USERS AND PROGRAM APPLICATIONS

In the past 2 years a large number of requests have been handled by TVA under a technology transfer contract with EPA. Listed below are those persons and organizations currently in possession of a tape copy of the program.

Accurex Corporation, Richard S. Merrill
Air Correction Divison, UOP, John Valente
American Natural Service Company, D. R. Tamir
Battelle Columbus Laboratories, Gerald Robinson
Black and Veatch, Don Cahalan
Burns & Roe, Inc., John Wysocki
Carnegie-Mellon University, Edward S. Rubin
Charles T. Main Consultants, Larry Decker
The City of New York, Environmental Protection Administration, M. Lieberstein
Combustion Engineering, B. M. Minor
Department of Energy, Larry Headley
Dravo Lime, David Hoffman
Exxon, John G. Stevens
Gibbs & Hill, Inc., Robert H. Boeckmann and K. J. Conrade
Gilbert Associates, David Becker
Intergroup Consulting Economists, Denis DePape
International Research & Technology Corporation, Anil Jagota
Kaiser Engineers, Inc., Douglas N. Trerice
Michael Baker, Inc., R. G. Knight
PEDCo, J. Scott Hartman
Radian Corporation, Jim Dickerman
Shell Development Company, Norman Samish
Stearns-Roger, Inc., Paul A. Ireland

A list of people who have been provided specific computer runs is shown below.

David Becker, Gilbert Associates
Sam Biondo, Department of Energy
J. J. Case, New York State Electric & Gas Corporation
Richard Chapman, Teknekron, Inc.
P. G. Cooper, Gulf Oil Research & Development
John Copeland, EPA
Wilson Cramer, U.S. Steel Corporation
Denis DePape, Intergroup Consulting Economists
John Eyraud, Energy Resource Company
Richard Furman, Florida Power & Light
Robert Gosik, EPA (Denver)
Robert Lane, Illinois Commerce Commission
Stephen J. Lutz, TRW, Inc.
Charles Martin, Indianapolis Power and Light

Jim Meyers, Detroit Edison
M. F. Patterson, Linde Division, Union Carbide Corporation
Robert Phillips, Radian Corporation
Dr. Joseph R. Polek, Catalytic, Inc.
Bill Reed, Carolina Power & Light
Dr. Thomas F. Rogers, Williams Brothers Engineering
Robert Statnick, EPA (Washington)
Dean Suagee, Bureau of Indian Affairs
D. R. Tamir, American Natural Service Company
Robin Watson, University of Tennessee

To date several uses of the program other than those for which it was intended have been tried. The program has been run simulating both industrial and utility boilers, smelter situations, partial scrubbing situations, plant optimization studies, and for comparisons of front-end coal-cleaning economics with total scrubbing. Probably the most important use to date has been support work for the National Economic Research Association's assessment of the effect of possible NSPS revisions on the electric utilities industry on a national scale. A similar study was done for EPA with Teknekron as the contractor. [EPA issued Federal Standards of Performance for New Stationary Sources (often called "new source performance standards" or NSPS).]

APPENDIX

INPUT DATA

EXAMPLE PRINTOUT

Example Case Inputs

```
1 1 1 1 1
1 1 1 1 1 1 1 1 1 1 1 1
1 1 1 1
INDUSTRIAL BRIEFING VST THICKENER-FILTER- OXIDATION
2 500 9000 10500 33 300 2 175 470 751
57.56 4.14 7.00 1.29 3.12 0.10 16.00 10.74 95 80 1 99.5 50.0
70 20 6 10.0 25 1 92 12 0 1.3 1 2 0.0 0.0 1500 3.0 2.85 5.0 0.0 0.0
15 40 2.20 40 2 95 1.5 70 1.2 7.0 0 10 20 14.7
3 0 0 0 50 .0000005 32 10 1.35 1 4 1 0.1
4 5.0 9999 3500 25 25 5280 1 12 2.50
9 16 5 10 8 8 10 8 3 50 10 14.9 1.17
8 50 12.50 2.0 0.12 0.029 17 262.5 209.7 1979 1980
2 .85 5 .8 .7 3 .6 1.0 1 1.1 1979 265
END
```

INDUSTRIAL BRIEFING VST THICKENER-FILTER, OXIDATION

```
                    *** INPUTS ***

BOILER CHARACTERISTICS
------ ---------------

MEGAWATTS =   500.

BOILER HEAT RATE =  9000. BTU/KWH

EXCESS AIR =  33. PERCENT, INCLUDING LEAKAGE

HOT GAS TEMPERATURE = 300. DEG F

COAL ANALYSIS, WT % AS FIRED :

  C       H      O      N      S      CL     ASH    H2O
57.56   4.14   7.00   1.29   3.12   0.10   16.00  10.74

SULFUR OVERHEAD =  95.0 PERCENT

ASH OVERHEAD =  80.0 PERCENT

HEATING VALUE OF COAL = 10500. BTU/LB

                              EFFICIENCY,   EMISSION,
FLYASH REMOVAL                    %         LBS/M BTU
------ -------                -----------   ---------

UPSTREAM OF SCRUBBER             99.5         0.06

WITHIN SCRUBBER                  50.0         0.03

COST OF UPSTREAM FLYASH REMOVAL EXCLUDED

ALKALI
------

LIMESTONE :

        CACO3         =  97.15 WT % DRY BASIS
        SOLUBLE MGO =    0.0
        INERTS        =   2.85
        MOISTURE CONTENT =   5.00 LB H2O/100 LBS DRY LIMESTONE
        LIMESTONE HARDNESS WORK INDEX FACTOR = 10.00
        LIMESTONE DEGREE OF GRIND FACTOR =  1.35

FLY ASH :

        SOLUBLE CAO =    0.0  WT %
        SOLUBLE MGO =    0.0
        INERTS      = 100.00


RAW MATERIAL HANDLING AREA
--- -------- -------- ----

NUMBER OF REDUNDANT ALKALI PREPARATION UNITS =     1
```

```
SCRUBBER SYSTEM VARIABLES
-------- ------ ---------

NUMBER OF OPERATING SCRUBBING TRAINS =   4

NUMBER OF REDUNDANT SCRUBBING TRAINS =   1

SPRAY TOWER LIQUID-TO-GAS RATIO =  70. GAL/1000 ACF

SPRAY TOWER GAS VELOCITY = 10.0 FT/SEC

VENTURI LIQUID-TO-GAS RATIO =  20. GAL/1000 ACF

VENTURI PRESSURE DROP = 10.0 IN H2O

SCRUBBER PRESSURE = 14.7 PSIA

STOICHIOMETRY =  1.30 MOLE CACO3 ADDED AS LIMESTONE
                      PER MOLE SO2 ABSORBED

ADIPIC ACID IN RECYCLE LIQUOR = 1500. PPM

ADIPIC ACID DECOMPOSITION RATIO =   3.

SO2 REMOVAL =  92. PERCENT

ENTRAINMENT LEVEL = 0.10 WT %

VHT RESIDENCE TIME =   6.0 MIN

EHT RESIDENCE TIME =  12.0 MIN

SO2 OXIDIZED IN SYSTEM =  95.0 PERCENT

AIR STOICHIOMETRY =  1.50 G-ATOM O /G-MOLE SO2 ABSORBED

SOLIDS IN RECIRCULATED SLURRY =  15.0 WT %


SOLIDS DISPOSAL SYSTEM
------ -------- ------

COST OF LAND =  3500.00 DOLLARS/ACRE

SOLIDS IN SYSTEM SLUDGE DISCHARGE =  70.0 WT %

SOLIDS IN CLARIFIER DISCHARGE = 40.0 WT %

CLARIFIER SOLIDS SETTLING RATE = 2.20 FT/HR

SOLIDS IN FILTER CAKE = 70.0 WT %

FILTRATION RATE = 1.20 TONS DRY SOLIDS/FT2 DAY
```

```
STEAM REHEATER (IN-LINE)
----- -------- ---------

SATURATED STEAM TEMPERATURE =  470. DEG F

HEAT OF VAPORIZATION OF STEAM =  751. BTU/LB

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC
```

```
INDUSTRIAL BRIEFING VST THICKENER-FILTER, OXIDATION

                    *** OUTPUTS ***

HOT GAS TO SCRUBBER
--- --- -- --------

          MOLE PERCENT      LB-MOLE/HR       LB/HR

CO2          12.315         0.2054E+05      0.9040E+06
HCL           0.007         0.1209E+02      0.4408E+03
SO2           0.238         0.3962E+03      0.2538E+05
O2            4.827         0.8051E+04      0.2576E+06
N2           73.869         0.1232E+06      0.3452E+07
H2O           8.745         0.1459E+05      0.2628E+06

SO2 CONCENTRATION IN SCRUBBER INLET GAS = 2375. PPM
                                        = 5.64 LBS / MILLION BTU

FLYASH EMISSION =  0.06 LBS/MILLION BTU
                =  0.030 GRAINS/SCF (WET)  OR    274. LB/HR

        SOLUBLE CAO IN FLY ASH =     0. LB/HR
        SOLUBLE MGO IN FLY ASH =     0.

HOT GAS FLOW RATE =  .1054E+07 SCFM (60 DEG F, 14.7 PSIA)
                  =  .1540E+07 ACFM (300. DEG F, 14.7 PSIA)

CORRESPONDING COAL FIRING RATE =  .4286E+06 LB/HR

HOT GAS HUMIDITY =  0.057 LB H2O/LB DRY GAS

WET BULB TEMPERATURE = 129. DEG F


WET GAS FROM SCRUBBER
--- --- ---- --------

          MOLE PERCENT      LB-MOLE/HR       LB/HR

CO2          11.748         0.2092E+05      0.9205E+06
SO2           0.018         0.3170E+02      0.2031E+04
O2            4.425         0.7878E+04      0.2521E+06
N2           69.203         0.1232E+06      0.3452E+07
H2O          14.606         0.2601E+05      0.4685E+06

SO2 CONCENTRATION IN SCRUBBER OUTLET GAS =  178. PPM

FLYASH EMISSION =  0.03 LBS/MILLION BTU
                =  0.014 GRAINS/SCF (WET)  OR    137. LB/HR

TOTAL WATER PICKUP =  422.  GPM
           INCLUDING   10.2 GPM ENTRAINMENT

WET GAS FLOW RATE =  .1125E+07 SCFM (60 DEG F, 14.7 PSIA)
                  =  .1274E+07 ACFM (129. DEG F, 14.7 PSIA)

WET GAS SATURATION HUMIDITY =  0.101 LB H2O/LB DRY GAS
```

```
FLUE GAS TO STACK
---- --- -- -----

             MOLE PERCENT       LB-MOLE/HR        LB/HR

CO2            11.730           0.2092E+05        0.9205E+06
SO2             0.018           0.3170E+02        0.2031E+04
O2              4.418           0.7878E+04        0.2521E+06
N2             69.093           0.1232E+06        0.3452E+07
H2O            14.742           0.2629E+05        0.4736E+06

SPECIFIED  SO2 REMOVAL EFFICIENCY =  92.0 %

CALCULATED SO2 EMISSION =   0.45 POUNDS PER MILLION BTU

CALCULATED SO2 CONCENTRATION IN STACK GAS =     178. PPM


FLYASH EMISSION =  0.03 LBS/MILLION BTU
                =  0.014 GRAINS/SCF (WET)  OR    137. LB/HR

STACK GAS FLOW RATE =  .1126E+07 SCFM (60 DEG F, 14.7 PSIA)
                    =  .1376E+07 ACFM (175. DEG F, 14.7 PSIA)


STEAM REHEATER (IN-LINE)
----- -------- ---------

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC

SQUARE PIPE PITCH = 2 TIMES ACTUAL PIPE O.D.

SATURATED STEAM TEMPERATURE =  470. DEG F

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

REQUIRED HEAT INPUT TO REHEATER = 0.6623E+08 BTU/HR

STEAM CONSUMPTION = 0.8818E+05 LBS/HR


                                     HEAT TRANSFER
OUTSIDE PIPE    PRESSURE DROP,       COEFFICIENT,
DIAMETER, IN.   IN. H2O              BTU/HR FT2 DEG F
  1.00               0.73              0.2079E+02


                REHEATER             NUMBER OF
                OUTSIDE PIPE         PIPES PER       NUMBER OF
                AREA, SQ FT          BANK PER        BANKS (ROWS)
                PER TRAIN            TRAIN           PER TRAIN

INCONEL         0.1198E+04              87              3
CORTEN          0.1311E+04              87              4
TOTAL           0.2509E+04              87              7
```

```
WATER BALANCE INPUTS
----- ------- ------


  RAINFALL(IN/YEAR)                 50.
  POND SEEPAGE(CM/SEC)*10**8        50.
  POND EVAPORATION(IN/YEAR)         32.



WATER BALANCE OUTPUTS
----- ------- -------


WATER AVAILABLE

  RAINFALL                           0. GPM                  66. LB/HR
  ALKALI                             5. GPM                2441. LB/HR
      TOTAL                          5. GPM                2507. LB/HR


WATER REQUIRED

  HUMIDIFICATION                   412. GPM              205730. LB/HR
  ENTRAINMENT                       10. GPM                5089. LB/HR
  DISPOSAL WATER                    62. GPM               30849. LB/HR
  HYDRATION WATER                   25. GPM               12634. LB/HR
  CLARIFIER EVAPORATION             29. GPM               14679. LB/HR
  POND EVAPORATION                   0. GPM                   0. LB/HR
  SEEPAGE                            0. GPM                   0. LB/HR

  TOTAL WATER REQUIRED             538. GPM              268982. LB/HR

NET WATER REQUIRED                 533. GPM              266475. LB/HR
```

```
SCRUBBER SYSTEM
-------- ------

TOTAL NUMBER OF SCRUBBING TRAINS (OPERATING+REDUNDANT) =  5

SO2 REMOVAL =  92.0 PERCENT

PARTICULATE REMOVAL IN SCRUBBER SYSTEM =  50.0 PERCENT

SPECIFIED  VENTURI PRESSURE DROP = 10.0 IN H2O

SPRAY TOWER PRESSURE DROP =  1.8 IN. H2O

TOTAL SYSTEM PRESSURE DROP = 18.4 IN. H2O

OVERRIDE TOTAL SYSTEM PRESSURE DROP = 20.0 IN. H2O

SPECIFIED   SPRAY TOWER L/G RATIO  =  70. GAL/1000 ACF

LIMESTONE ADDITION = 0.4882E+05 LB/HR DRY LIMESTONE

SPECIFIED  LIMESTONE STOICHIOMETRY  =  1.30 MOLE CACO3 ADDED AS LIMESTONE
                                                PER MOLE SO2 ABSORBED

SOLUBLE CAO FROM FLY ASH =  0.0  MOLE PER MOLE SO2 ABSORBED

TOTAL SOLUBLE MGO        =  0.0  MOLE PER MOLE SO2 ABSORBED

TOTAL STOICHIOMETRY      =  1.30 MOLE SOLUBLE (CA+MG)
                                 PER MOLE SO2 ABSORBED

SCRUBBER INLET LIQUOR PH =  5.94

MAKE UP WATER =  533. GPM

MAKE-UP ADIPIC ACID = 0.1423E+03 LB/HR

OXIDATION AIR RATE = 0.3779E+05LB/HR
                   = 0.8231E+04 SCFM (60 DEG F,1 ATM)

CROSS-SECTIONAL AREA PER SCRUBBER =   531. SQ FT
```

```
SOLIDS DISPOSAL SYSTEM
------ -------- ------

TOTAL CLARIFIER(S) CROSS-SECTIONAL AREA =    2230. SQ FT

SYSTEM SLUDGE DISCHARGE
------ ------ ---------
                                                    SOLID     LIQUID
                                                    COMP,     COMP,
SPECIES              LB-MOLE/HR       LB/HR          WT %      PPM

CASO3 .1/2 H2O       0.1819E+02      0.2348E+04      3.18
CASO4 .2H2O          0.3461E+03      0.5956E+05     80.73
CACO3                0.1033E+03      0.1034E+05     14.02
INSOLUBLES           ----------      0.1529E+04      2.07
H2O                  0.1712E+04      0.3085E+05
CA++                 0.6433E+01      0.2578E+03                8154.
MG++                 0.0             0.0                          0.
SO3--                0.4802E-01      0.3845E+01                 122.
SO4--                0.3145E+00      0.3021E+02                 955.
CL-                  0.1214E+02      0.4305E+03               13614.
AD=                  0.3249E+00      0.4743E+02                1500.


TOTAL DISCHARGE FLOW RATE = 0.1054E+06 LB/HR
                          =    120.    GPM

TOTAL DISSOLVED SOLIDS IN DISCHARGE LIQUID =  24346. PPM

DISCHARGE LIQUID PH =  7.14
```

```
SCRUBBER SLURRY BLEED
-------- ------ -----

SPECIES             LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O      0.1819E+02    0.2348E+04
CASO4 .2H2O         0.3461E+03    0.5956E+05
CACO3               0.1033E+03    0.1034E+05
INSOLUBLES          ----------    0.1529E+04
H2O                 0.2264E+05    0.4079E+06
CA++                0.8506E+02    0.3409E+04
MG++                0.0           0.0
SO3--               0.6350E+00    0.5084E+02
SO4--               0.4158E+01    0.3994E+03
CL-                 0.1606E+03    0.5692E+04
AD=                 0.4295E+01    0.6271E+03

TOTAL FLOW RATE = 0.4919E+06 LB/HR
                =     893.    GPM

TOTAL SUPERNATE RETURN
----- --------- ------

SPECIES             LB-MOLE/HR    LB/HR

H2O                 0.2014E+05    0.3628E+06
CA++                0.7565E+02    0.3032E+04
MG++                0.0           0.0
SO3--               0.5648E+00    0.4522E+02
SO4--               0.3698E+01    0.3553E+03
CL-                 0.1428E+03    0.5062E+04
AD=                 0.3820E+01    0.5578E+03

TOTAL FLOW RATE = 0.3718E+06 LB/HR
                =     744.    GPM

SUPERNATE TO WET BALL MILL
--------- -- --- ---- ----

SPECIES             LB-MOLE/HR    LB/HR

H2O                 0.1654E+04    0.2981E+05
CA++                0.6215E+01    0.2491E+03
MG++                0.0           0.0
SO3--               0.4640E-01    0.3715E+01
SO4--               0.3038E+00    0.2919E+02
CL-                 0.1173E+02    0.4159E+03
AD=                 0.3139E+00    0.4582E+02

TOTAL FLOW RATE = 0.3055E+05 LB/HR
                =      61.    GPM
```

```
LIMESTONE SLURRY FEED
--------- ------ ----

SPECIES              LB-MOLE/HR    LB/HR

CACO3                0.4738E+03    0.4743E+05
SOLUBLE MGO          0.0           0.0
INSOLUBLES           ----------    0.1391E+04
H2O                  0.1763E+04    0.3176E+05
CA++                 0.6215E+01    0.2491E+03
MG++                 0.0           0.0
SO3--                0.4640E-01    0.3715E+01
SO4--                0.3038E+00    0.2919E+02
CL-                  0.1173E+02    0.4159E+03
AD=                  0.1288E+01    0.1881E+03

TOTAL FLOW RATE =  0.8137E+05  LB/HR
                =     103.     GPM

SUPERNATE RETURN TO SCRUBBER OR EHT
--------- ------ -- -------- -- ---

SPECIES              LB-MOLE/HR    LB/HR

H2O                  0.1848E+05    0.3330E+06
CA++                 0.6943E+02    0.2783E+04
MG++                 0.0           0.0
SO3--                0.5184E+00    0.4150E+02
SO4--                0.3395E+01    0.3261E+03
CL-                  0.1311E+03    0.4647E+04
AD=                  0.3506E+01    0.5119E+03

TOTAL FLOW RATE =  0.3413E+06  LB/HR
                =     683.     GPM

RECYCLE SLURRY TO SPRAY TOWER
------- ------ -- ----- -----

SPECIES              LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O       0.1815E+04    0.2344E+06
CASO4 .2H2O          0.3455E+05    0.5946E+07
CACO3                0.1031E+05    0.1032E+07
INSOLUBLES           ----------    0.1526E+06
H2O                  0.2260E+07    0.4072E+08
CA++                 0.8491E+04    0.3403E+06
MG++                 0.0           0.0
SO3--                0.6339E+02    0.5075E+04
SO4--                0.4151E+03    0.3988E+05
CL-                  0.1603E+05    0.5682E+06
AD=                  0.4288E+03    0.6261E+05

TOTAL FLOW RATE =  0.4910E+08  LB/HR
                =   89182.     GPM
```

```
RECYCLE SLURRY TO VENTURI
------- ------ -- -------

SPECIES             LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O      0.5187E+03    0.6697E+05
CASO4 .2H2O         0.9872E+04    0.1699E+07
CACO3               0.2947E+04    0.2950E+06
INSOLUBLES          ----------    0.4360E+05
H2O                 0.6458E+06    0.1163E+08
CA++                0.2426E+04    0.9724E+05
MG++                0.0           0.0
SO3--               0.1811E+02    0.1450E+04
SO4--               0.1186E+03    0.1139E+05
CL-                 0.4580E+04    0.1624E+06
AD=                 0.1225E+03    0.1789E+05

TOTAL FLOW RATE = 0.1403E+08 LB/HR
                =  25481.     GPM

CLARIFIER UNDERFLOW SLURRY
--------- --------- ------

SPECIES             LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O      0.1819E+02    0.2348E+04
CASO4 .2H2O         0.3461E+03    0.5956E+05
CACO3               0.1033E+03    0.1034E+05
INSOLUBLES          ----------    0.1529E+04
H2O                 0.5993E+04    0.1080E+06
CA++                0.2251E+02    0.9024E+03
MG++                0.0           0.0
SO3--               0.1681E+00    0.1346E+02
SO4--               0.1101E+01    0.1057E+03
CL-                 0.4250E+02    0.1507E+04
AD=                 0.1137E+01    0.1660E+03

TOTAL FLOW RATE = 0.1844E+06 LB/HR
                =    278.     GPM

SUPERNATE FROM CLARIFIER
--------- ---- ---------

SPECIES             LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O      0.0           0.0
CASO4 .2H2O         0.0           0.0
CACO3               0.0           0.0
INSOLUBLES          ----------    0.0
H2O                 0.1586E+05    0.2857E+06
CA++                0.5957E+02    0.2387E+04
MG++                0.0           0.0
SO3--               0.4447E+00    0.3560E+02
SO4--               0.2912E+01    0.2797E+03
CL-                 0.1124E+03    0.3986E+04
AD=                 0.3008E+01    0.4392E+03

TOTAL FLOW RATE = 0.2928E+06 LB/HR
                =    586.     GPM
```

```
FILTER CAKE SLURRY
------ ---- ------
SPECIES              LB-MOLE/HR     LB/HR

CASO3 .1/2 H2O       0.1819E+02     0.2348E+04
CASO4 .2H2O          0.3461E+03     0.5956E+05
CACO3                0.1033E+03     0.1034E+05
INSOLUBLES           ----------     0.1529E+04
H2O                  0.1712E+04     0.3085E+05
CA++                 0.6433E+01     0.2578E+03
MG++                 0.0            0.0
SO3--                0.4802E-01     0.3845E+01
SO4--                0.3145E+00     0.3021E+02
CL-                  0.1214E+02     0.4305E+03
AD=                  0.3249E+00     0.4743E+02

TOTAL FLOW RATE = 0.1054E+06 LB/HR
                =    120.    GPM

FILTRATE FROM FILTER
-------- ---- ------
SPECIES              LB-MOLE/HR     LB/HR

CASO3 .1/2 H2O       0.0            0.0
CASO4 .2H2O          0.0            0.0
CACO3                0.0            0.0
INSOLUBLES           ----------     0.0
H2O                  0.4281E+04     0.7712E+05
CA++                 0.1608E+02     0.6446E+03
MG++                 0.0            0.0
SO3--                0.1201E+00     0.9612E+01
SO4--                0.7862E+00     0.7553E+02
CL-                  0.3036E+02     0.1076E+04
AD=                  0.8121E+00     0.1186E+03

TOTAL FLOW RATE = 0.7905E+05 LB/HR
                =    158.    GPM
```

```
           INDUSTRIAL BRIEFING VST POND,                                            CASE 001

                              PARTICULATE REMOVAL INVESTMENT AND OPERATING COST
                                                      PARTICULATE EMISSION REGULATION (LB ASH/MILLION BTU):   0.06
WPSUL CONTENT (%):            3.12                    FLUE GAS TEMPERATURE (COLD) (F):                       300.0
ASH CONTENT (%):              16.00                   FLUE GAS TEMPERATURE (HOT) (F):                        700.0
BTU RATING:                   10500                   COST OF ELECTRICITY ($/KWHR):                            0.03
BOILER TYPE:                  DRY PULVERIZED COAL     COST OF STEAM ($/THOUSAND LB):                           2.00
NO. OF SCRUBBERS:                 4                   FIRST YEAR CAPITAL CHARGE FACTOR:                        0.194
SCRUBBER VELOCITY (FT/M):     600.0                   BAGHOUSE RATIO (OPER. SQ.FT./ACTUAL SQ.FT.):             0.80
PLANT SIZE (MW):                500                   BAG COST ($/SQ.FT.):                                     0.70
OPERATING HRS/YR:              4512                   BAG LIFE(YEARS):                                        3.00
PUMPING RATE (GAL/1000 ACF):  20.00                   FLUE GAS REHEAT TEMPERATURE (F):                        175.
SCA RATIO:                    1.100                   CHEMICAL ENGINEERING PLANT INDEX:                       265.0
  (ACTUAL SQ.FT./CALC. SQ.FT.)



                                   ELECTROSTATIC PRECIPITATORS

                                         COLD            HOT          BAGHOUSE FABRIC FILTERS      SCRUBBERS

REQUIRED REMOVAL EFFICIENCY (%):         99.50          99.50                   99.50               99.50
DRIFT VELOCITY (FT/M):                   25.78          20.00
SPECIFIC COLLECTION AREA (SQ.FT./ACFM):  226.05         291.41
COLLECTION AREA (SQ.FT.):             348077.3       684896.3                721033.4
TOTAL CORONA POWER (KW):                 467.4          713.4
AUXILIARY POWER (KW):                    293.0          620.5                   461.5
FAN POWER (KW):                          241.0          367.8                  1204.9              8998.3
PUMP POWER (KW):                                                                                    3796.0
TOTAL POWER (KW):                       1001.4         1701.8                  1666.3             12794.3
OPERATING AIR/CLOTH RATIO:                                                        2.7
INSTALLED AIR/CLOTH RATIO:                                                        2.1
REQUIRED PRESSURE DROP (INCHES):           1.0            1.0                     5.0                37.3
DIAMETER (FEET):                                                                                       32
REQUIRED REHEAT (BTU/HR):                                                                      58660416.0
STEAM SUPPLY/YR (THOUSAND LB):                                                                   294084.2
INSTALLED COST (1979 DOLLARS):      $  5046359     $  8923927              $ 13998391          $ 21061472
FIRST YEAR CAPITALIZED COST:        $   979162     $  1731539              $  2716154          $  4086627
ANNUAL POWER COST:                  $   131027     $   222672              $   218038          $  1674110
ANNUAL OPERATING AND
 MAINTENANCE COST (1979 DOLLARS):   $    95696     $   139421              $    70773          $  1105205
REPLACEMENT COST (1979 DOLLARS):                                           $   220712
ANNUAL REHEAT COST:                                                                            $   588168



TOTAL ANNUAL COST:                  $  1205885     $  2093632              $  3225677          $  7454110
ANNUALIZED COST OF POWER(MILLS/KWHR):     0.53           0.93                    1.43                3.30
```

RAW MATERIAL HANDLING AND PREPARATION
INCLUDING 2 OPERATING AND 1 SPARE PREPARATION UNITS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
CAR SHAKER AND HOIST	20HP SHAKER 7.5HP HOIST	1	28582.	1866.
CAR PULLER	25HP PULLER, 5HP RETURN	1	49345.	1866.
UNLOADING HOPPER	16FT DIA, 10FT STRAIGHT SIDE HT, CS	1	4180.	7711.
UNLOADING VIBRATING FEEDER	3.5HP	1	12134.	1866.
UNLOADING BELT CONVEYOR	20FT HORIZONTAL, 5HP	1	17527.	0.
UNLOADING INCLINE BELT CONVEYOR	310FT, 50HP	1	60670.	24875.
UNLOADING PIT DUST COLLECTOR	POLYPROPYLENE BAGTYPE, 2200 CFM,7.5HP	1	5258.	12438.
UNLOADING PIT SUMP PUMP	60GPM, 70FT HEAD, 5HP	1	3371.	746.
STORAGE BELT CONVEYOR	200FT, 5HP	1	57974.	16169.
STORAGE CONVEYOR TRIPPER	30FPM, 1HP	1	13482.	2488.
MOBILE EQUIPMENT	SCRAPPER TRACTOR	1	136171.	0.
RECLAIM HOPPER	7FT WIDE, 4.25FT HT, 2FT WIDE BOTTOM, CS	2	1079.	1741.
RECLAIM VIBRATING FEEDER	3.5HP	2	24268.	3731.
RECLAIM BELT CONVEYOR	200FT, 5HP	1	40447.	8706.
RECLAIM INCLINE BELT CONVEYOR	193FT, 40HP	1	37750.	13930.
RECLAIM PIT DUST COLLECTOR	POLYPROPYLENE BAG TYPE	1	5258.	12438.
RECLAIM PIT SUMP PUMP	60GPM, 70FT HEAD, 5HP	1	3371.	746.
RECLAIM BUCKET ELEVATOR	90FT HIGH, 75HP	1	80894.	1617.
FEED BELT CONVEYOR	60.FT HORIZONTAL 7.5HP	1	20223.	1368.
FEED CONVEYOR TRIPPER	30 FPM, 1HP	1	13482.	2488.
FEED BIN	13FT DIA, 21FT STRAIGHT SIDE HT, COVERED, CS	3	16179.	29851.
BIN WEIGH FEEDER	14FT PULLEY CENTERS, 2HP	3	54603.	3731.
GYRATORY CRUSHERS	75HP	3	159765.	15672.
L MILL DUST COLLECTORS	POLYPROPYLENE BAG TYPE 2200 CFM, 7.5HP	3	15774.	37313.

BALL MILL	12.2TPH, 165.HP	3	475169.	43835.
MILLS PRODUCT TANK	5500 GAL 10FT DIA, 10FT HT, FLAKEGLASS LINED CS	3	14561.	22761.
MILLS PRODUCT TANK AGITATOR	10HP	3	24673.	1119.
MILLS PRODUCT TANK SLURRY PUMP	51.GPM, 60FT HEAD, 2.HP, 2 OPERATING AND 1 SPARES	3	8661.	2612.
ADIPIC ACID ADD STORAGE SILO	1992.FT3, 11.9FT DIA 27.8FT HT 60 DEG CONE	1	13596.	7972.
PNEUMATIC CONVEYOR SYSTEM	10. HP	1	9101.	5099.
ADDITION FEED BIN	RUBBER LINED	1	3514.	1602.
SCREW FEEDER	30 FT LONG, 6 IN D, SS	1	4988.	498.
ADDITIVE DUST COLLECTOR	POLYPROPLENE BAG TYPE 450 CFM, 1.5 HP	1	3505.	249.
SLURRY FEED TANK	54242.GAL, 21.0FT DIA, 21.0FT HT, FLAKEGLASS-LINED CS	1	20187.	15001.
SLURRY FEED TANK AGITATOR	47.HP	1	43570.	3216.
SLURRY FEED TANK PUMPS	26.GPM, 60 FT HEAD, 1.HP, 4 OPERATING AND 4 SPARE	8	22334.	6965.
			---------	---------
TOTAL EQUIPMENT COST			1505640.	314285.

SCRUBBING

INCLUDING 4 OPERATING AND 1 SPARE SCRUBBING TRAINS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
F.D. FANS	20.0IN H2O, WITH 1615. HP MOTOR AND DRIVE	5	1873839.	113586.
VENTURI		5	578888.	69736.
SHELL			933048.	
RUBBER LINING			1394698.	
MIST ELIMINATOR			464438.	
SLURRY HEADER AND NOZZLES			574586.	
TOTAL SPRAY SCRUBBER COSTS		5	3366768.	323889.
REHEATERS		5	1908890.	149806.
SOOTBLOWERS		75	357280.	211441.
VENTURI-OXIDATION HOLD TANK	73496.GAL 18.4FT DIA, 36.9FT HT, FLAKEGLASS-LINED CS	5	142171.	105739.
VENTURI-OXIDATION TANK AGITATO	52.HP	5	180218.	66502.
VENTURI-OXIDATION PUMPS	6370.GPM 100 FT HEAD, 295.HP 4OPERATING AND 6 SPARE	10	431969.	59925.
EFFLUENT HOLD TANK	294302.GAL, 36.9FT DIA, 36.9FT HT, FLAKEGLASS-LINED CS	5	357079.	259529.
EFFLUENT HOLD TANK AGITATOR	74.HP	5	480979.	177486.
ABSORBER RECYCLE PUMPS	11148.GPM, 100FT HEAD, 516.HP, 8 OPERATING AND 7 SPARE	15	992108.	114793.
OXIDATION BLEED PUMPS	893.GPM, 60 FT HEAD 25.HP, 1 OPERATING AND 1 SPARE	2	15365.	5079.
OXIDATION AIR BLOWER	2058.SCFM, 206.HP	6	164606.	4478.
OXIDATION SPARGER	9.2 FT DIA RING	5	38013.	37002.
MAKEUP WATER PUMPS	3185.GPM, 200.FT HEAD, 268.HP, 1 OPERATING AND 1 SPARE	2	33084.	3357.
			---------	---------
TAL EQUIPMENT COST			10921254.	1702341.

WASTE DISPOSAL

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
ABSORBER BLEED RECEIVING TANK	73496.GAL, 18.4FT DIA, 36.9FT HT, FLAKGLASS-LINED CS	1	28434.	21148.
ABSORBER BLEED TANK AGITATOR	42.HP	1	32699.	2413.
THICKENER FEED PUMP	893.GPM, 60FT HEAD, 25.HP, 1 OPERATING AND 1 SPARE	2	15365.	5079.
THICKENER	2230.SQ.FT., 53.FT DIA, 5.6TANK FT HT	1	91998.	58228.
THICKENER UNDERFLOW SLURRY PUMPS	278.GPM, 9.2FT HEAD, 1.HP , 1 OPERATING AND 1 SPARE	2	8132.	3051.
THICKENER OVERFLOW PUMPS	586.GPM, 75.0FT HEAD, 19.HP, 1 OPERATING AND 1 SPARE	2	11049.	1121.
THICKENER OVERFLOW TANK	9666.GAL, 17.1FT DIA, 5.6FT HT	1	3532.	2172.
FILTER FEED TANK	4591.GAL, 9.2FT DIA, 9.2FT HT, FLAKEGLASS-LINED CS	1	3892.	2892.
FILTER FEED TANK AGITATOR	7.HP	1	5761.	425.
FILTER FEED SLURRY PUMP	139.GPM, 50FT HEAD, 4.HP, 2 OPERATING AND 1 SPARE	3	12050.	3380.
FILTER	369.SQ FT FILTRATION AREA	2	265829.	43420.
FILTRATE PUMP (PER FILTER)	79.GPM, 20.0FT HEAD, 1.HP, 2 OPERATING AND 2 SPARE	4	18185.	1845.
FILTRATE SURGE TANK	2610.GAL, 7.6FT DIA, 7.6FT HT	2	9665.	981.
FILTRATE SURGE TANK PUMP	158.GPM, 85.0FT HEAD, 6.HP, 2 OPERATING AND 2SPARE	1	1558.	958.
TOTAL EQUIPMENT COST			508150.	147114.

LIMESTONE SLURRY PROCESS -- BASIS: 500 MW UNIT, 1980 STARTUP

PROJECTED CAPITAL INVESTMENT REQUIREMENTS - INDUSTRIAL BRIEFING VST THICKENER-FILTER, OXIDATION CASE 002

	INVESTMENT, THOUSANDS OF 1979 DOLLARS				DISTRIBUTION
	RAW MATERIAL HANDLING AND PREPARATION	SCRUBBING	WASTE DISPOSAL	TOTAL	PERCENT OF DIRECT INVESTMENT
EQUIPMENT					
MATERIAL	1506.	10921.	508.	12935.	38.1
LABOR	314.	1702.	147.	2164.	6.4
PIPING					
MATERIAL	232.	3168.	386.	3786.	11.1
LABOR	93.	858.	142.	1092.	3.2
DUCTWORK					
MATERIAL	0.	2952.	0.	2952.	8.7
LABOR	0.	2247.	0.	2247.	6.6
FOUNDATIONS					
MATERIAL	131.	216.	22.	369.	1.1
LABOR	540.	423.	46.	1009.	3.0
POND CONSTRUCTION	0.	0.	0.	0.	0.0
STRUCTURAL					
MATERIAL	278.	554.	0.	832.	2.4
LABOR	114.	869.	0.	982.	2.9
ELECTRICAL					
MATERIAL	175.	689.	182.	1045.	3.1
LABOR	347.	1152.	453.	1953.	5.7
INSTRUMENTATION					
MATERIAL	154.	928.	50.	1133.	3.3
LABOR	17.	116.	62.	195.	0.6
BUILDINGS					
MATERIAL	62.	32.	39.	133.	0.4
LABOR	58.	30.	37.	129.	0.4
SERVICES AND MISCELLANEOUS	126.	841.	65.	1032.	3.0
SUBTOTAL DIRECT INVESTMENT	4147.	27698.	2141.	33985.	100.0
ENGINEERING DESIGN AND SUPERVISION	373.	2493.	193.	3059.	9.0
CONSTRUCTION EXPENSES	663.	4432.	342.	5438.	16.0
CONTRACTOR FEES	207.	1385.	107.	1699.	5.0
CONTINGENCY	415.	2770.	214.	3399.	10.0
SUBTOTAL FIXED INVESTMENT	5806.	38777.	2997.	47579.	140.0
ALLOWANCE FOR STARTUP AND MODIFICATIONS	464.	3102.	240.	3806.	11.2
INTEREST DURING CONSTRUCTION	464.	3102.	240.	3806.	11.2
SUBTOTAL CAPITAL INVESTMENT	6734.	44981.	3476.	55192.	162.4
LAND	7.	3.	3.	13.	0.0
WORKING CAPITAL	177.	1183.	91.	1452.	4.3
TOTAL CAPITAL INVESTMENT	6919.	46167.	3570.	56656.	166.7

```
LIMESTONE SLURRY PROCESS -- BASIS:  500 MW UNIT, 1980 STARTUP

PROJECTED REVENUE REQUIREMENTS - INDUSTRIAL BRIEFING VST THICKENER-FILTER, OXIDATION                    CASE 002

                                    DISPLAY SHEET FOR YEAR=     1
                                    ANNUAL OPERATION KW-HR/KW =  4512

                       36.89 TONS PER HOUR                         DRY                    SLUDGE
                              TOTAL CAPITAL INVESTMENT        56656000                          TOTAL
                                                                                                ANNUAL
                                             ANNUAL QUANTITY         UNIT COST,$                COST,$
                                             ---------------         -----------                ------
      DIRECT COSTS
      ------------
        RAW MATERIAL
        ------------
          LIMESTONE                            110.1 K TONS           8.00/TON                  881100
                                                                                             ----------
             SUBTOTAL RAW MATERIAL                                                              881100

        CONVERSION COSTS
        ----------------
          OPERATING LABOR AND
            SUPERVISION                      46110.0 MAN-HR          12.50/MAN-HR               576400
            UTILITIES
            STEAM                           397890.0 K LB             2.00/K LB                 795800
            PROCESS WATER                   144340.0 K GAL            0.12/K GAL                 17300
            ELECTRICITY                   48587910.0 KWH              0.03/KWH                 1409000
          MAINTENANCE
            LABOR AND MATERIAL                                                                 2089000
          ANALYSES                            4020.0 HR              17.00/HR                    68400
                                                                                             ----------
             SUBTOTAL CONVERSION COSTS                                                         4955900

             SUBTOTAL DIRECT COSTS                                                             5837000

      INDIRECT COSTS
      --------------
        DEPRECIATION                                                                           1839700
        COST OF CAPITAL AND TAXES, 14.90% OF UNDEPRECIATED INVESTMENT                          8441800
        INSURANCE & INTERIM REPLACEMENTS,  1.17% OF TOTAL CAPITAL INVESTMENT                    662900
        OVERHEAD
          PLANT,  50.0% OF CONVERSION COSTS LESS UTILITIES                                     1366900
          ADMINISTRATIVE, RESEARCH, AND SERVICE,
            10.0% OF OPERATING LABOR AND SUPERVISION                                             57600
                                                                                             ----------
             SUBTOTAL INDIRECT COSTS                                                          12368900

             SUBTOTAL ANNUAL REVENUE REQUIREMENT                                              18205900

             SLUDGE FIXATION COSTS          166400.0 TONS             5.00/TON                  832000
                                                                                             ----------
             TOTAL ANNUAL REVENUE REQUIREMENT                                                 19037900
                                                                                             ----------
             EQUIVALENT UNIT REVENUE REQUIREMENT, MILLS/KWH                                       8.44
----------------------------------------------------------------------------------------------------------------------
         HEAT RATE   9000. BTU/KWH    -    HEAT VALUE OF COAL    10500 BTU/LB    -    COAL RATE    966900 TONS/YR
```

LIMESTONE SLURRY PROCESS -- BASIS: 500 MW UNIT, 1980 STARTUP

PROJECTED LIFETIME REVENUE REQUIREMENTS - INDUSTRIAL BRIEFING VST THICKENER-FILTER, OXIDATION CASE 002

TOTAL CAPITAL INVESTMENT: $ 56656000

YEARS AFTER POWER UNIT START	ANNUAL OPERATION, KW-HR /KW	POWER UNIT HEAT REQUIREMENT, MILLION BTU /YEAR	POWER UNIT FUEL CONSUMPTION, TONS COAL /YEAR	SULFUR REMOVED BY POLLUTION CONTROL PROCESS, TONS/YEAR	BYPRODUCT RATE, EQUIVALENT TONS/YEAR DRY SLUDGE	SLUDGE FIXATION FEE $/TON DRY SLUDGE	ADJUSTED GROSS ANNUAL REVENUE REQUIREMENT EXCLUDING SLUDGE FIXATION COST, $/YEAR	TOTAL ANNUAL SLUDGE FIXATION COST, $/YEAR	NET ANNUAL INCREASE IN TOTAL REVENUE REQUIREMENT, $	CUMULATIVE NET INCREASE IN TOTAL REVENUE REQUIREMENT, $
1	4512	20304000	966900	26300	166400	5.00	18205900	832000	19037900	19037900
2	4643	20893500	994900	27100	171300	5.00	18092700	856500	18949200	37987100
3	4775	21487500	1023200	27900	176100	5.00	17979600	880500	18860100	56847200
4	4906	22077000	1051300	28600	181000	5.00	17864300	905000	18769300	75616500
5	5037	22666500	1079400	29400	185800	5.00	17748600	929000	18677600	94294100
6	5169	23260500	1107600	30200	190700	5.00	17633100	953500	18586600	112880700
7	5300	23850000	1135700	30900	195500	5.00	17516100	977500	18493600	131374300
8	5432	24444000	1164000	31700	200400	5.00	17399500	1002000	18401500	149775800
9	5563	25033500	1192100	32500	205200	5.00	17281200	1026000	18307200	168083000
10	5694	25623000	1220100	33200	210000	5.00	17162200	1050000	18212200	186295200
11	5695	25627500	1220400	33200	210100	5.00	16889200	1050500	17939700	204234900
12	5695	25627500	1220400	33200	210100	5.00	16615100	1050500	17665600	221900500
13	5695	25627500	1220400	33200	210100	5.00	16341000	1050500	17391500	239292000
14	5695	25627500	1220400	33200	210100	5.00	16066900	1050500	17117400	256409400
15	5695	25627500	1220400	33200	210100	5.00	15792800	1050500	16843300	273252700
16	5537	24916500	1186500	32300	204300	5.00	15331500	1021500	16353000	289605700
17	5379	24205500	1152600	31400	198400	5.00	14869200	992000	15861200	305466900
18	5221	23494500	1118800	30500	192600	5.00	14406300	963000	15369300	320836200
19	5064	22788000	1085100	29600	186800	5.00	13943400	934000	14877400	335713600
20	4906	22077000	1051300	28600	181000	5.00	13478500	905000	14383500	350097100
21	4748	21366000	1017400	27700	175100	5.00	13012500	875500	13888000	363985100
22	4591	20659500	983800	26800	169400	5.00	12546600	847000	13393600	377378700
23	4433	19948500	949900	25900	163500	5.00	12078200	817500	12895700	390274400
24	4275	19237500	916100	25000	157700	5.00	11608700	788500	12397200	402671600
25	4118	18531000	882400	24000	151900	5.00	11139100	759500	11898600	414570200
26	3960	17820000	848600	23100	146100	5.00	10666700	730500	11397200	425967400
27	3802	17109000	814700	22200	140300	5.00	10193000	701500	10894500	436861900
28	3645	16402500	781100	21300	134500	5.00	9719100	672500	10391600	447253500
29	3487	15691500	747200	20400	128600	5.00	9242100	643000	9885100	457138600
30	3329	14980500	713400	19400	122800	5.00	8763300	614000	9377300	466515900
TOT	146001	657004500	31286100	852000	5385900		439586400	26929500	466515900	
LIFETIME AVERAGE INCREASE IN UNIT REVENUE REQUIREMENT										
DOLLARS PER TON OF COAL BURNED							14.05	0.86	14.91	
MILLS PER KILOWATT-HOUR							6.02	0.37	6.39	
CENTS PER MILLION BTU HEAT INPUT							66.91	4.10	71.01	
DOLLARS PER TON OF SULFUR REMOVED							515.95	31.60	547.55	
REVENUE REQUIREMENT DISCOUNTED AT 10.0% TO INITIAL YEAR, DOLLARS							156636400	8783200	165419600	
LEVELIZED INCREASE IN UNIT REVENUE REQUIREMENT EQUIVALENT TO DISCOUNTED REQUIREMENT OVER LIFE OF POWER UNIT										
DOLLARS PER TON OF COAL BURNED							15.35	0.86	16.21	
MILLS PER KILOWATT-HOUR							6.58	0.37	6.95	
CENTS PER MILLION BTU HEAT INPUT							73.10	4.09	77.19	
DOLLARS PER TON OF SULFUR REMOVED							563.85	31.61	595.46	

INVESTIGATION OF ALKALI FEEDRATE AUTOMATIC CONTROL FOR THE SLURRY PROCESS OF LIMESTONE SCRUBBERS

Patrick Garrett
University of Cincinnati

I. ABSTRACT

Three automatic control methods are analyzed for dense limestone feedrate to the recirculating slurry based upon mathematical modeling, simulation studies, and experimental pilot scrubber tests: (1) stoichiometric control which is based on meaningful material balance measurements, but lacking in its visibility of influential slurry factors such as the process gain and flyash alkaline content which can result in significant limestone feedrate errors; (2) slurry pH measurement and control which is responsive to an agglomerate of slurry reactions invisible to stoichiometric control, but requires a rigorous design for stable operation due to the potential interaction of the slurry and control system gain and phase shift elements; and (3) combined stoichiometric and pH control which utilizes the most independent process measurements to determine limestone feedrate, and is optimum in minimizing the disorder of the slurry to load changes or other process upsets.

II. INTRODUCTION

Early attempts at the optimization of limestone scrubbers involved combinations of L/G ratio, holding tank residence time, and recirculating solids which would maximize SO_2 removal while minimizing the risks of internal scaling. However, an incomplete understanding of observed effects conspired to prevent achievement of the desired control. More recently the characterization and operating experience with these processes has reached a level of maturity to better support a productive effort into the optimization of their control.

The application of automatic control to environmental processes including limestone-alkali wet scrubbers is especially promising because of its potential for increased reliability, economy of operation, and consistent reduction in the variance of controlled variables with varying operating conditions and operator proficiency. Implementation of solutions to the major obstacles to limestone scrubbing, the latter including internal scaling and insufficient SO_2 removal, in practice can be optimized only through the disorder reducing capabilities of process automation.

The effectiveness of limestone scrubbing is primarily dependent upon the ability to accommodate and regulate the manifold factors associated with the recirculating slurry. This problem is all the more a challenge because of the limited number of useful slurry process measurements and actuator inputs available, and the economic considerations associated with the energy and feedstock requirements of full-scale scrubbers. These consequences prompted the utilization of the rigorous methods of process modeling and computer simulation to obtain the results reported here.

III. CONCLUSIONS

Effective control of the recirculating slurry is essential for successful operation of limestone scrubbers at the reliability, SO_2 removal level, and efficiency of interest. Three control methods have been analyzed and specified for the addition of dense limestone to the slurry, which constitutes the most difficult scrubber control action to realize effectively because of enumerated competing slurry factors.

Stoichiometric control of limestone addition is widespread in application, unconditionally stable, and meaningfully based on slurry material balance measurements. A recommended design for stoichiometric control is presented. However, analysis shows this control method more accurately to be dead reckoning based on feedforward measurements, and lacking in its visibility of influential slurry factors which can result in significant errors in the limestone feedrate. Therefore, its utility as the sole determiner of limestone addition is compromised and not recommended.

PH control of the limestone feedrate is geared to solve for many of the influential slurry factors which are invisible to stoichiometric control, because the pH measurement represents an agglomerate of the slurry reactions. However, the feedback implementation of this control method requires a rigorous design for stable and responsive performance due to the interaction of the slurry and controller frequency-determining elements. Nevertheless, pH control of limestone addition can be adequate, but not optimum, as the sole determiner of limestone feedrate. The effectiveness of pH control was also found to be maximized for operation within a narrow range of slurry pH setpoints about 5.5. A detailed design is provided which has been proofed in a described pilot scrubber test.

The combination of stoichiometric and pH control utilizes the most independent process measurements to determine the limestone feedrate, and is optimum in the sense that further reduction in the slurry process disorder is available than either method is capable of individually. The deadband pH and self-optimizing adaptive stoichiometric ratio envelope control additions necessary to realize improved performance result in increased system complexity, but can be implemented with conventional process controllers.

IV. STOICHIOMETRIC CONTROL

Absorption of SO_2 by a limestone slurry in a wet scrubbing process is typically followed by subsequent oxidation of the calcium sulfite to calcium sulfate as described by equations (1) and (2). If the scrubber physical chemistry is straightforward, the stoichiometric ratio of limestone to sulfur corresponding to the desired sulfur removal can be calculated and the slurry dense limestone addition operated with stoichiometric control. In practice, limestone is added at a rate proportional to the mass inlet SO_2 rate which is determined by a setpoint equated to the desired stoichiometric ratio as shown in Figure 1. It should be noted that this stoichiometric ratio setpoint is, in fact, only a feedforward gain adjustment which is prevented from agreeing with the true slurry stoichiometry by the primary factors of entrained flyash alkaline content, limestone quality and solubility, and chloride content of the slurry. Nevertheless, feedforward control is useful in this application because of its ability to correct for SO_2 load disturbances to the sluggish buffered limestone slurry. However, its utility as the sole determiner of limestone addition is compromised because measured inlet SO_2 does not reflect slurry sulfite solubility and other factors which affect reliable and consistent scrubber operation.

$$CaO + SO_2 \rightarrow CaSO_3 \qquad (1)$$

$$CaSO_3 + \tfrac{1}{2}O_2 \rightarrow CaSO_4 \qquad (2)$$

Feedforward limestone addition systems referenced to stoichiometric control are based upon the material balance of moles of SO_2 absorbed to moles of Ca added. Real-time computation of these parameters can be provided from the scrubber flue gas inlet duct by means of a dedicated computation unit using temperature, pressure, and SO_2 analyzer measurements defined by equations (3) through (7). The SO_2 inlet concentration has subtracted from it the allowable scrubber SO_{2s} exit concentration setpoint yielding a ΔSO_2 input to the mass flow computer. Calculation of the slurry limestone addition volumetric flow rate F is then based upon neutralization of the required ΔSO_2 at the anticipated stoichiometric ratio (SR). The ΔP signal is typically provided from a scrubber inlet-duct venturi, and the only manual operation required is a periodic wet chemistry orsat determination of the total flue gas molecular weight Mw.

$$Q = A \cdot k \cdot \sqrt{\frac{\Delta P \cdot T}{P \cdot Mw}} \cdot \frac{P \cdot T_{std}}{P_A \cdot T} \ m^3/\text{minute} \qquad (3)$$

$$\Delta SO_2 = SO_{2_i} - SO_{2_s} \ \text{ppm} \qquad (4)$$

$$M = Q \cdot \Delta SO_2 \cdot Mw_{SO_2} \ \text{kg/minute} \qquad (5)$$

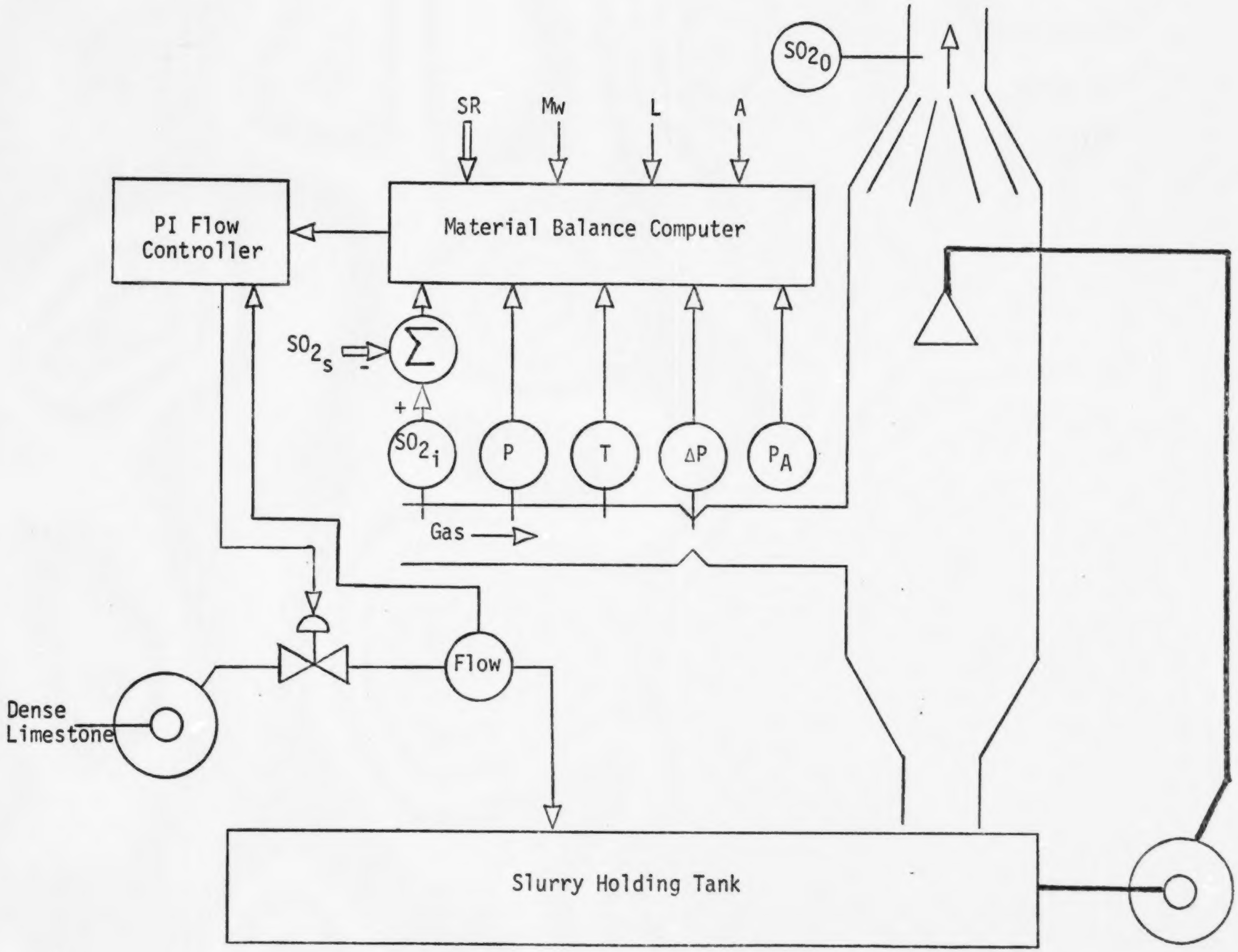

SO_{2_0}
SR
Mw
L
A
PI Flow Controller
Material Balance Computer
SO_{2_s}
-
+
Σ
SO_{2_i}
P
T
ΔP
P_A
Gas
Flow
Dense Limestone
Slurry Holding Tank

Figure 1. Feedforward Stoichiometric Limestone Addition Control System

$$L = \text{Limestone\% purity} \cdot \text{kg/liter weight} \cdot \text{density} \cdot \text{specific gravity} \qquad (6)$$

$$F = \frac{M \cdot SR}{L} \text{ liters/minute} \qquad (7)$$

With reference to Figure 1, five measured variables and four settable coefficients are input to the feedforward stoichiometric material balance computer. The output dense limestone volumetric flow rate F provides the remote setpoint input signal to a proportional plus integral limestone flow controller. This mechanization provides for more accurate limestone addition, by overcoming disturbances in the limestone addition system such as pump cavitation, than the direct application of the flow signal to the valve. Rigorous analysis of this control system shows that it is not truly feedforward because of the existence of the SO_{2_S} exit target setpoint in the material balance equations. True feedforward control requires the substitution of outlet SO_{2_O} for SO_{2_S} to eliminate the error at all times when $SO_{2_O} \neq SO_{2_S}$. This error is representative of the limitations of feedforward control which is completely dependent upon the accuracy to which it models the process, but is inavoidable to force SO_2 removal toward the required exit target SO_{2_S} value.

Parameter	Description	Parameter	Description
A	Duct area	M_W	Gas molecular wt
P	Duct pressure	ΔSO_2	SO_2 absorbed
Tstd	Standard temperature	$M_{W_{SO_2}}$	SO_2 molecular wt
P_A	Atmospheric pressure	L	Limestone weight/liter
T	Duct temperature	k	$\frac{\text{g/gmole} \cdot \text{mmHg}}{°\text{k} \cdot \text{mmH}_2\text{O}}$
ΔP	Differential pressure		
		SR	Stoke ratio$\cdot\frac{\text{mole } CaCO_3}{\text{mole } SO_2}$

Table 1. Material Balance Computer Parameters

V. PH CONTROL

Feedback control of limestone addition by means of a slurry pH measurement offers the capability for accommodating the important factors listed in Table 2 which feedforward stoichiometric control cannot correct for. However, the control system design is substantially complicated with feedback control because the slurry process involving the limestone dissolution and holding tank neutralization reactions is now positioned between the limestone addition and pH measurement as shown in Figure 2. Consequently, a detailed analysis of the slurry process is required and the determination of a compatible controller which can achieve and maintain a slurry pH setpoint with unconditional stability. The design procedure applied was to develop a valid mathematical model which describes the system, generate a computer simulation of the control system with this model, evaluate its dynamics and stability with likely controller mechanizations under both steady-state and load-disturbance conditions, and proof the final control system design by evaluating it on an actual scrubber process.

process gain
flyash alkaline effect
calcium sulfite scaling
%SO_2 removal
limestone variability and solubility
slurry chloride effect
throwaway sludge

Table 2. pH Related Slurry Factors

The major obstacles to the effectiveness of limestone scrubbers have been internal scaling and insufficient SO_2 removal. The ability of pH control to influence and regulate the limestone slurry factors presented in Table 2 results from the use of a slurry pH measurement, which is an indicator of slurry chemical reactions, and an appropriate control law. Flyash alkali, limestone alkalinity, and chloride content from coals and make-up water containing HCl all affect the slurry alkali utilization and can be regulated under pH control. This is especially important because of the relationship of alkali utilization to calcium sulfite solubility and the effect of the latter on operating reliability due to internal scaling. For example scaling is normally prevented at a slurry pH of about 5.5 and below, which produces an alkali utilization of 85% and above. However, lower slurry pH values produce reduced SO_2 removal resulting in a narrow range of practical slurry pH operating points about 5.5 which provide a consistent baseline of SO_2 removal generally in the 70% range.

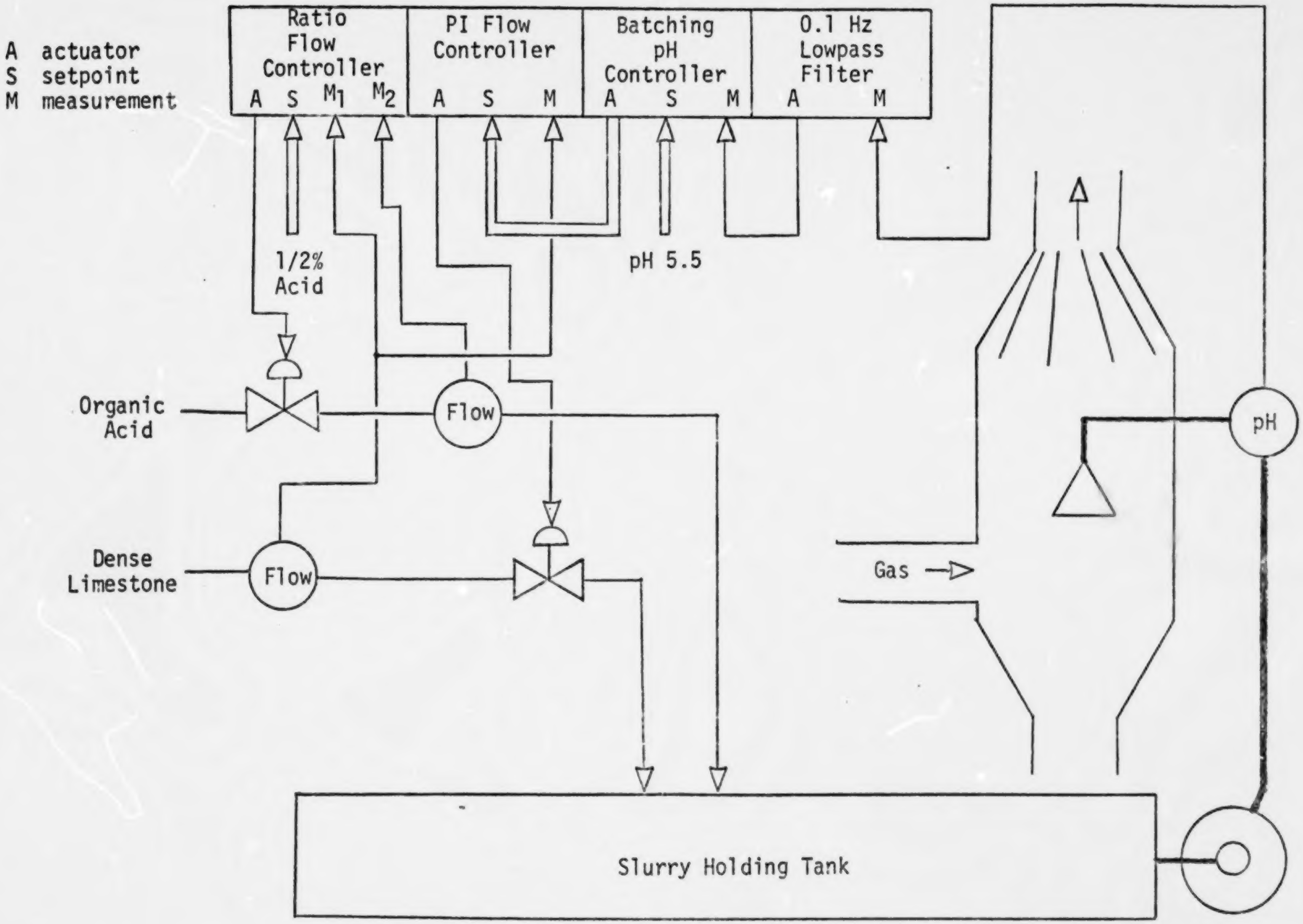

A actuator
S setpoint
M measurement
Ratio Flow Controller
A S M_1 M_2
PI Flow Controller
A S M
Batching pH Controller
A S M
0.1 Hz Lowpass Filter
A M
1/2% Acid
pH 5.5
Organic Acid
Flow
Dense Limestone
Flow
pH
Gas
Slurry Holding Tank

Figure 2. pH Feedback Limestone Addition Control System

With sulfite solubility maintained and scaling prevented with pH control, SO_2 removal can be boosted to the 90% range by the ratio-control addition of organic acid buffer agents such as adipic acid also included in Figure 2. This flow controller adds buffer to the slurry in proportion to limestone feedrate. In addition, a high alkali utilization factor offers improved controllability of the slurry process because of an increased neutralization process gain (ΔpH/Δ limestone feedrate), and reduces the amount of undissolved limestone in the sludge. Both controllability and sludge minimization are further enhanced with forced slurry aeration due, respectively, to CO_2 stripping accelerating limestone dissolution and oxidation of the calcium sulfite to calcium sulfate.

A. Modeling And Simulation Results

A mathematical model of limestone slurry dissolution and neutralization reactions was developed and validated based on a regression fit to the Shawnee TCA pilot plant limestone scrubber depletion runs. This model consists of three elements in cascade defined by Table 3 and shown in Figure 4, where t_d is the slurry holding tank residence time. The kinetics of limestone dissolution are difficult to model, but significant when control methods are considered. It is notable that this function G_D is identically a geometric progression whereby an infinite number of recirculations are required for complete dissolution. The neutralization time constant for a back-mixed reactor slurry tank is also essentially the residence time t_d, and becomes converted to a pure integrator function as a result of its position in the slurry recirculation circuit. An interesting characteristic of limestone dissolution G_N is its resonance phenomenon illustrated in Figure 3 which arises from the distributed nature of this batching process.

An EAI TR-20 analog computer was utilized for simulation studies of the slurry model, and was instrumental in providing insight into the dynamics and control requirements for this scrubber process. Figure 4 shows a composite of the process and controller mechanization simulated including both feedforward and Smith predictor compensators. A PI controller option included was an antiwindup batching feature which disconnects the error signal to the integrator for output actuation commands of <0%>100%. This prevents the controller from deleteriously integrating the error signal further when its actuation capability has limit cycled. The control system with these compensators was simulated with step changes in load of ± 0.2 pH unit, which correspond to ± 1000 ppm changes in SO_2 load, about a steady-state scrubber load of 0.5 pH representing 2500 ppm SO_2. For all runs the slurry pH setpoint was maintained at 5.5, resulting in the conversion of the incremental slurry model into an absolute model for computer simulation purposes.

Element	Characteristic	Function	Shawnee TCA
Dissolution G_D	Geometric	$\frac{k_1}{1 - k_2 e^{-t_d s}}$	$\frac{0.037 - (0.037)(0.063)^n}{1 - 0.963}$
Process Gain k_p	Nonlinear	$\frac{\Delta pH}{\Delta \text{ limestone feed}}$	0.045
Neutralization G_N	Integrator	$\frac{1/t_d s + 1}{1 - 1/t_d s + 1}$	$\frac{1}{10s}$
Transfer Function	Closed Loop	$\frac{-\frac{s}{t_d}(1 - k_2 e^{-t_d s})}{s^2(1-k_2 e^{-t_d s}) + \frac{k \cdot k_1 \cdot k_p}{t_r \cdot t_d}(t_r s+1)}$	

Table 3. Summary of Slurry Mathematical Model

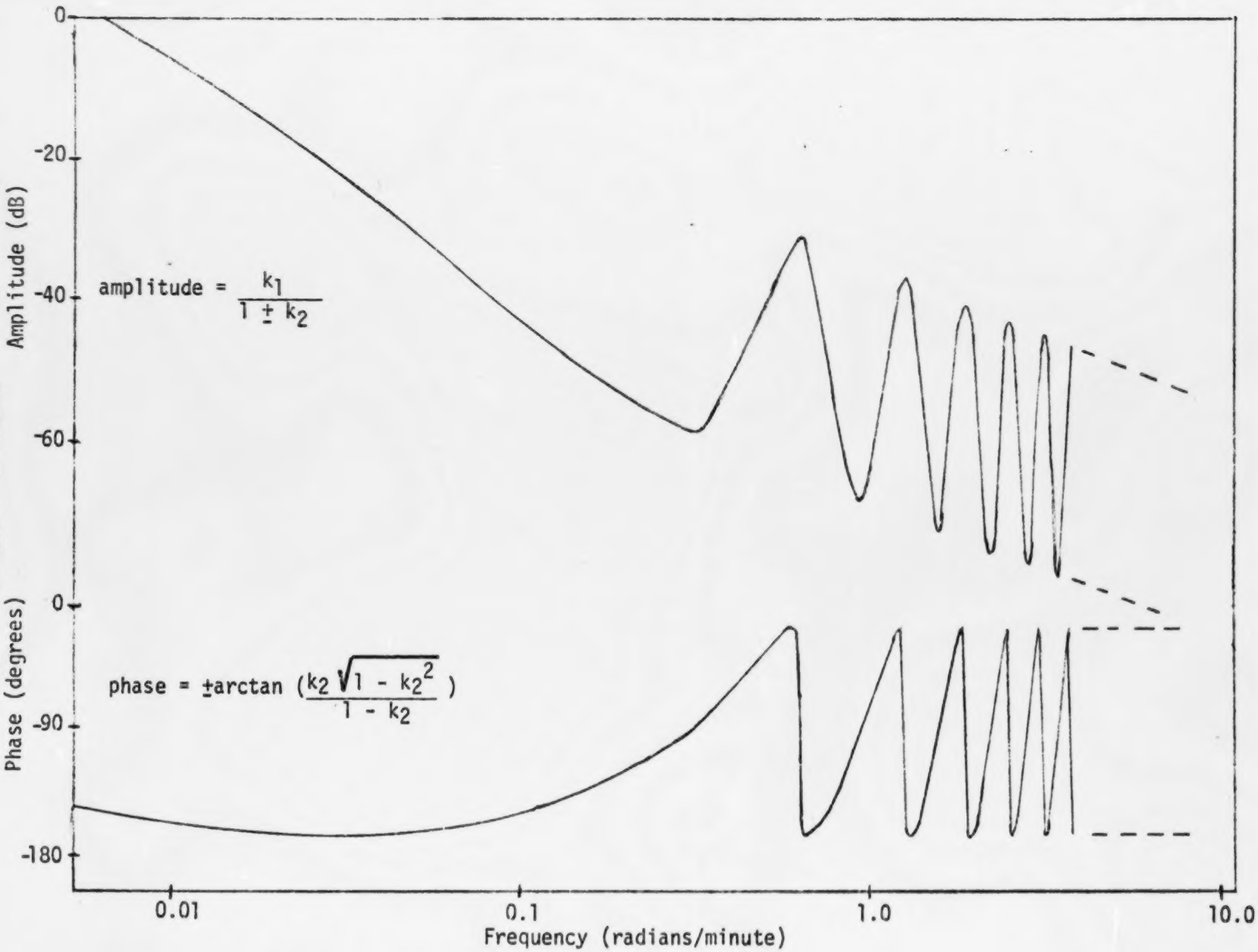


Figure 3. Limestone Dissolution Resonance Phenomenon

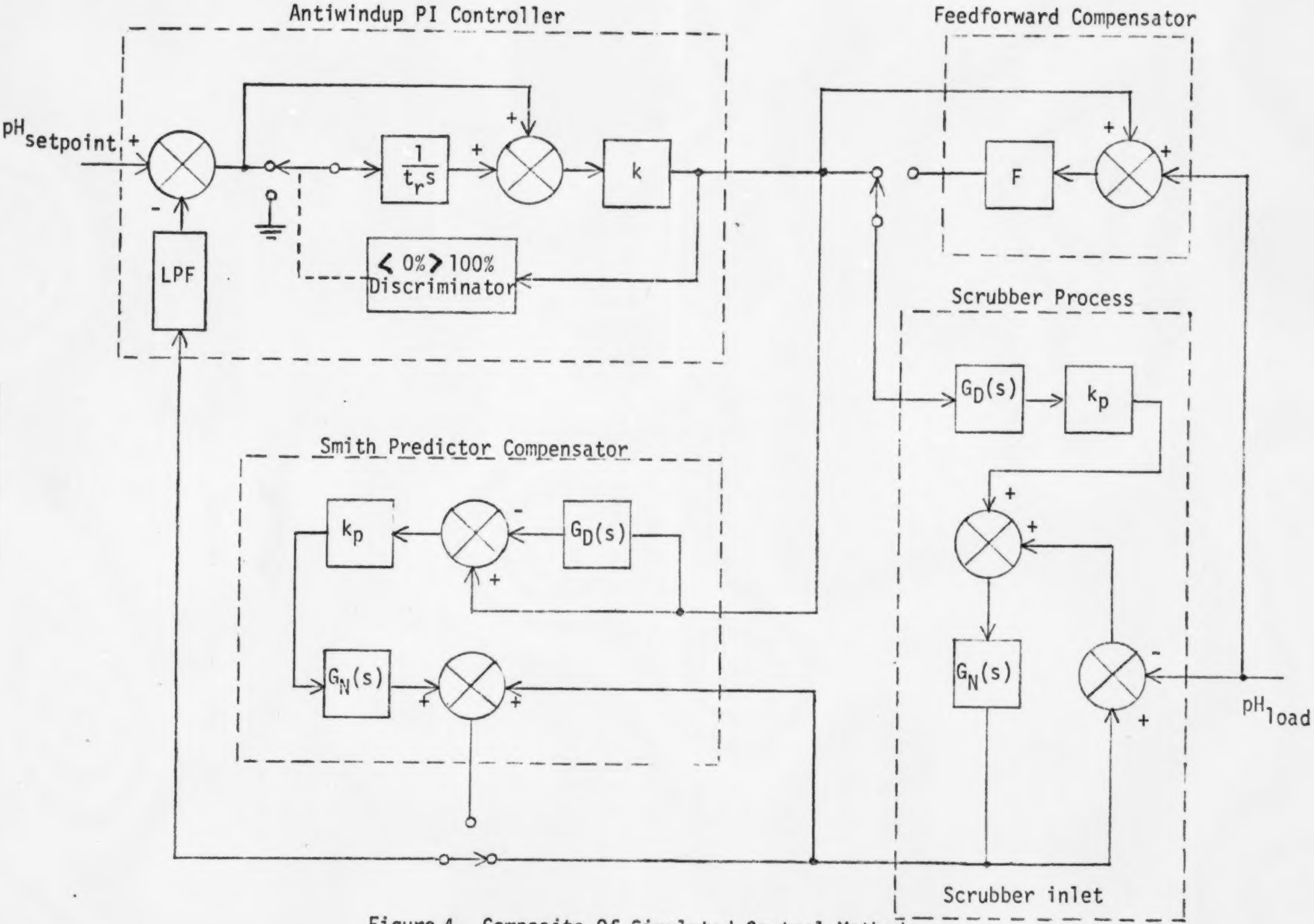

Antiwindup PI Controller
Feedforward Compensator
$pH_{setpoint}$
+
-
LPF
$\frac{1}{t_r s}$
+
+
k
< 0% > 100% Discriminator
F
+
+
Scrubber Process
$G_D(s)$
k_p
+
+
$G_N(s)$
-
+
pH_{load}
Scrubber inlet
Smith Predictor Compensator
k_p
-
$G_D(s)$
+
$G_N(s)$
+
+

Figure 4. Composite Of Simulated Control Methods

The pH control simulation results are summarized by Figure 5, and show the shortest time for recovery to the operating setpoint following load changes is provided by the combination of feedforward and feedback control of Run (a). This reflects the ability of feedforward control to respond to inlet changes in SO_2 a priori to influencing the slurry, with pH feedback control geared to solve for a posteriori slurry unknowns such as flyash alkali content. The Smith predictor of Run (b) did not provide the anticipated control improvement with its compensatory algorithm, which utilizes the process deadtime information t_d to make more intelligent corrections over that possible with the delayed process error signal alone. The reason is evident from Figure 4, showing the lag elements of both the neutralization integrator $1/t_d s$ and controller integrator $1/t_r s$ inavoidably in the path of the load change information on its circuit to the Smith predictor input. This effectively defeats the purpose of the Smith predictor; i.e., to transform the visibility of the process to the controller to appear as if the process has no time delay.

Runs (c) and (d) describe the process response with straight pH feedback, which was determined to be capable of accurate and well-behaved control. Based on the criteria of deviation from setpoint following a load disturbance and recovery time to setpoint, the controller antiwindup feature was found to offer substantial advanjage over a conventional proportional plus integral controller. Although load increases require approximately the same recovery time with and without antiwindup, the more troublesome load decreases representing excess limestone entrainment in the slurry recovered to setpoint in a notably shorter interval. This was a result of discontinued controller integration of the error signal during intervals when the controller output has limit cycled.

B. Experimental Evaluation

Translation of the modeling, simulation, and analysis of pH control of limestone feedrate into a control system design resulted in the mechanization shown in Figure 2. A batching controller with adjustable antiwindup limits was obtained from Foxboro Corporation, to insure compatibility with the existing scrubber control components, and connected in cascade with one of their conventional PI controllers. Each controller has its own measurement input, but only the pH batching controller has an independent setpoint and only the secondary limestone flow controller has an output to the process. This configuration was implemented because of the effectiveness of the secondary flow loop in compensating for disturbances in sluggish processes, such as the limestone slurry, to ameliorate their influence on the primary pH controlled variable. A narrow range 4.5 - 6.5 pH probe was specified and supplied by Electrofact, Model 135 G/Z, with mechanical probe wiping action and a buffer-solution pressurized ref-

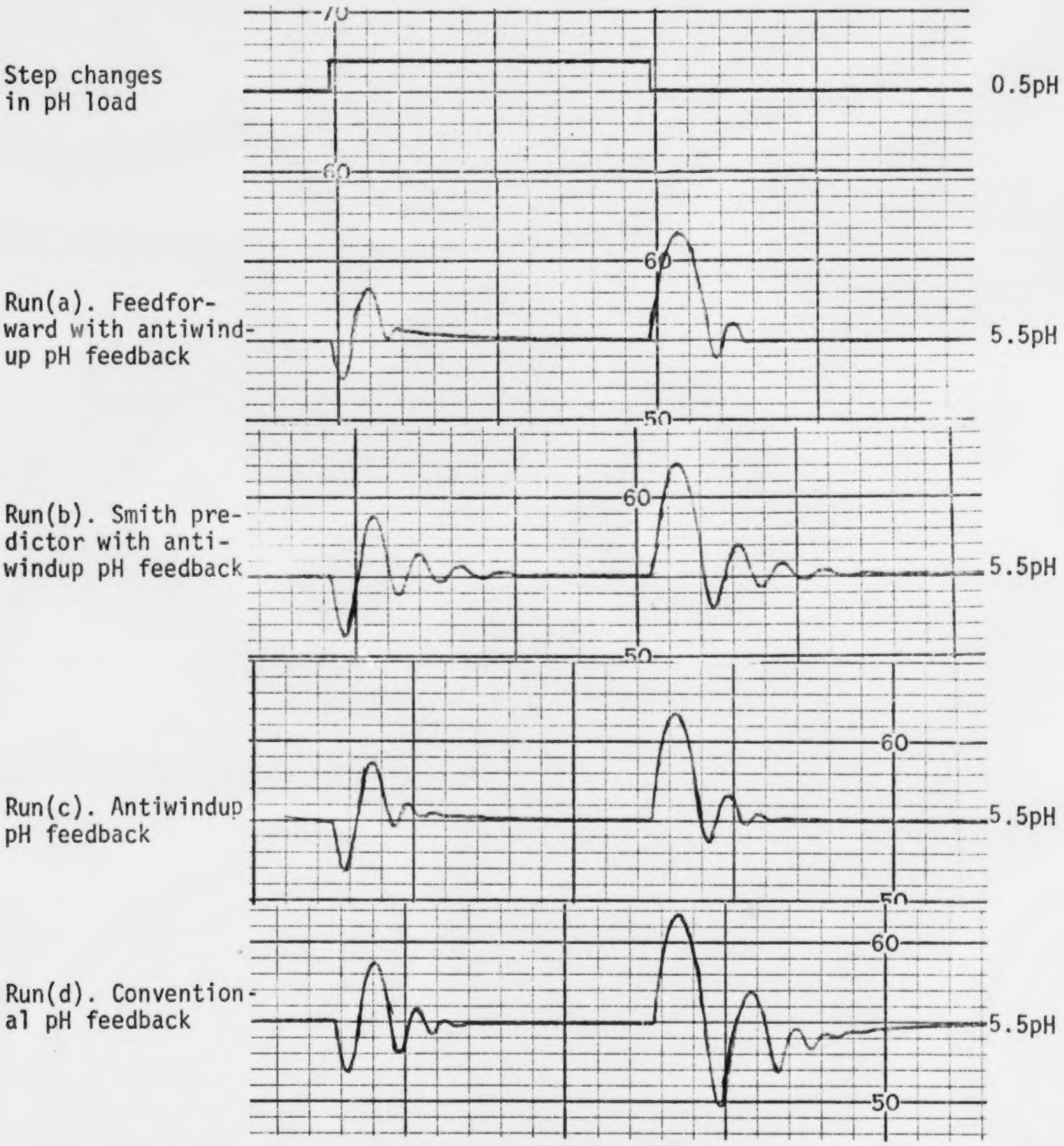


Figure 5. pH Control Simulation Runs

Legend: 100 minutes/horizontal division
0.1pH/vertical division

erence electrode for immersion in the difficult slurry environment. A custom 4-pole sharp-cutoff Butterworth lowpass filter was also included in the pH signal circuit to prevent process noise and transients from entering the control system thereby insuring smooth, well-behaved control dynamics.

Following installation and debugging of this control system on the TVA Shawnee TCA limestone pilot scrubber near Paducah, Kentucky, without the organic acid buffer addition system, planned experiments were conducted from July 26th to the 30th, 1979. Preliminary activities included careful online tuning of the limestone flow and pH controllers, obtained from observation of the subsidence of the controlled pH and limestone flow variables following oscillation induced by excessive controller gain. Table 4 lists the controller final tuning adjustments, where the gain of the flow controller is six times that of the pH controller and satisfies the cascade control requirement that the secondary (flow) loop have at least four times the gain of the primary (pH) loop.

The neutralization process gain kp was experimentally determined at the three indicated slurry pH setpoints of 5.3, 5.55, and 5.8 as a component of the test plan and with the aid of equation (8). The pH controller was placed in manual and 5.6 liters/minute of dense limestone added in each instance to the specific limestone feedrate, and the change in ΔpH recorded as shown in Table 5. A comparison of laboratory and measured data disclosed a variance in pH during the first day of the test series. This was determined to be a low calibration error with the Electrofact pH instrument averaging 0.46 pH and was corrected in situ. Consequently, the first and second process gain measurements were obtained at an indicated slurry pH of 5.3 and 5.8, but in fact were at 5.76 and 6.0, respectively. The third was obtained at the correctly indicated pH value of 5.55. The neutralization gain was found to have a value of 0.08 over this slurry pH operating range, and the inlet SO_2 variation during these tests was within the experimental test accuracy. This gain value was obtained with a 4.1 minute holding tank.

Controller	Proportional Band	Reset Rate
pH	200%	5 minutes
flow	30%	3 minutes

Table 4. pH Controller Tuning Adjustments

Time	1700/7-29	1100/7-26	1600/7-26
pH	5.55	5.76	6.0
SO_{2inlet}	1600 ppm	2040 ppm	1880 ppm
Δliters/minute	5.6	5.6	5.6
ΔpH	0.45	0.56	0.2
k_p	0.08	0.1	0.035

Table 5. Neutralization Process Gain Data

$$\text{process gain kp} = \frac{\Delta \text{ pH}}{\Delta \text{limestone feedrate}} \qquad (8)$$

Since the loop-gain around a closed-loop control system cannot exceed unity for unconditional stability, neither may the gain product of the cascaded controllers with the slurry process. Equation (9) determines the loop gain to be a responsive but stable 0.6, as determined by the pH controller proportional band, its scaling, and the process gain. The closed-loop gain of the secondary flow loop is 1.0 because its mechanization effectively relays the primary controller output signal. Figure 6 presents a segment of the pH and limestone feedrate controlled variables recorded during July 28th and early the 29th during factorial runs to determine scrubber performance under pH control. Between 6PM and 9PM scrubber operating conditions were sufficiently consistent to permit both limestone feedrate and pH to line out at a pH setpoint of 5.82. This verifies the ability to achieve a variance of the pH controlled variable to within ±0.01 pH unit, and illustrates the performance possible with continuous linear control. Successful pH control was obtained even with a restricted limestone feedrate pump range of 2 - 12 liters/minute due to reported pump control and pipe restriction problems. This is evident in the non-symmetrical limestone flow half-cycling between midday and 4PM in Figure 6.

In conclusion, an independent analysis[1] of these factorial runs under pH control indicates that pH and SO_2 removal correlate with a regression coefficient of 0.82, which increases to 0.86 when

---

[1] "Operations of Shawnee TCA with Automated pH Control," GCA Corporation report to U.S. Environmental Protection Agency, IERL-PMB, RTP, NC, August 21, 1979.

Reproduced from
best available copy.

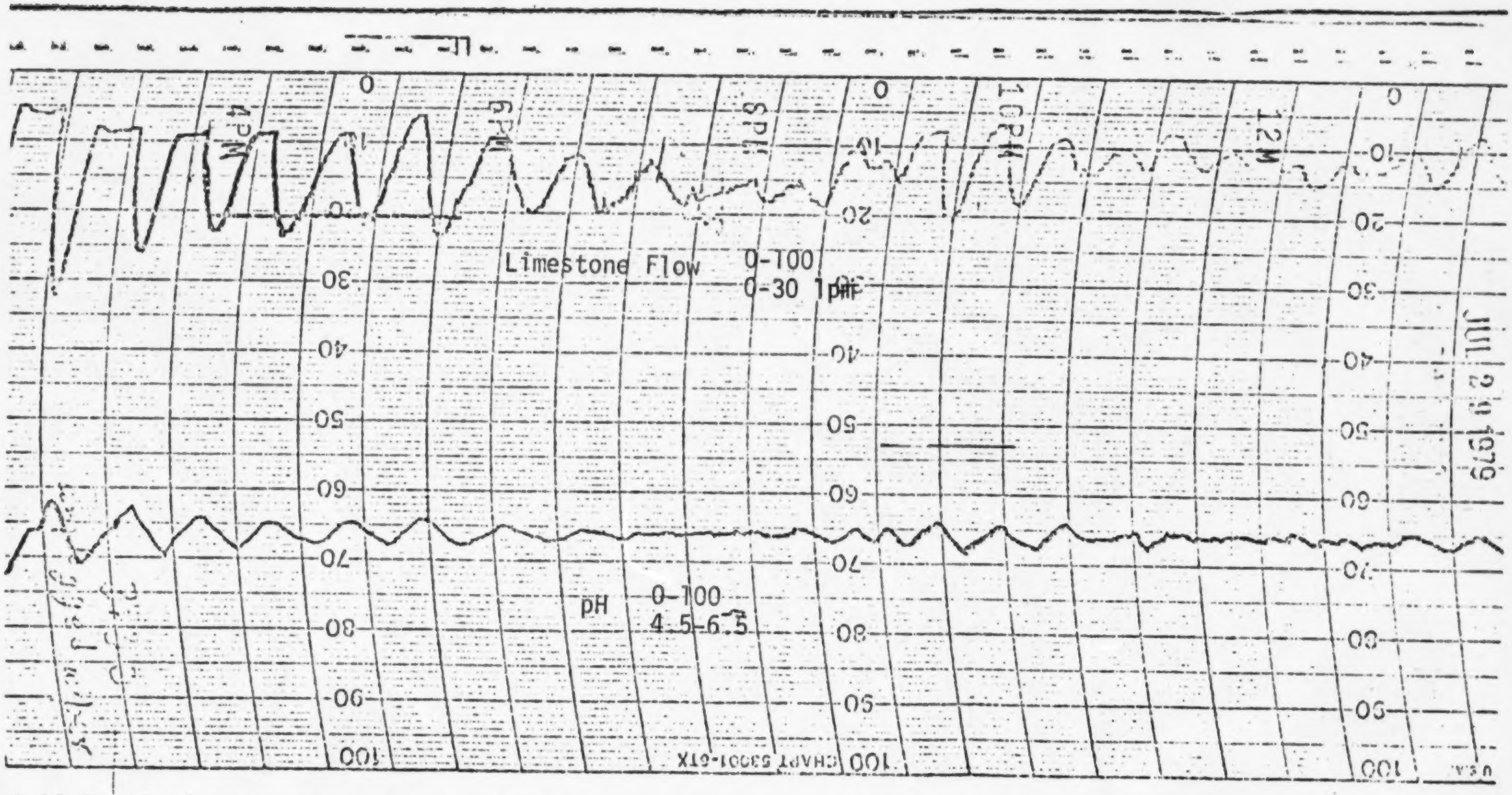


Figure 6. Limestone Flow And pH Controlled Variables

inlet SO_2 is included. This additionally validates the utility of pH control, and substantiates the value of a combination of pH and stoichiometric control of the limestone feedrate.

$$\text{loop gain} = (\text{pH controller})(\text{flow controller})(\text{neutralization gain}) \qquad (9)$$

$$= \left(\frac{100\%}{200\%} \cdot \frac{30.3 \text{ liters}}{2\text{pH}}\right)(1.0)\left(0.08 \frac{\Delta\text{pH}}{\frac{\Delta\text{liters}}{\text{minute}}}\right)$$

$$= 0.6$$

VI. COMBINED STOICHIOMETRIC AND pH CONTROL

Results from the independent simulation study of pH control with feedforward compensation plus the regression analysis of SO_2 removal correlation with pH for the Shawnee pH control test both indicate improved regulation of the limestone slurry, even with major load disturbances, is possible when inlet SO_2 is also utilized in the control of limestone addition. The appropriate combination of both feedforward stoichiometric and feedback pH control offers an optimum control scheme for the limestone addition system. Implementation involves the hybrid combination of the control system designs represented by Figures 1 and 2, with modifications, and capitalizes on their respective process a priori material balance and a posteriori slurry reaction pH information to reduce disorder in the slurry process below that possible with stoichiometric or pH control separately.

Figure 7 describes the parallel combination of these two control methods with both outputs identically scaled for the full delivery capability of the dense limestone addition system. Parallel operation is advantageous primarily for stability considerations, which are minimized when the least number of frequency determining elements such as controller integrators are placed in series. In operation a slow integrating-only batching pH controller automatically adjusts the stoichiometric ratio setpoint to minimize the feedforward control system error by integrating the offset between the slurry pH and its setpoint, where its batching limits are adjustable to delimit the range of stoichiometric gain change. However, slurry pH excursions outside a preselected deadband, for an example deadband between 5.4 - 5.7 pH, reflect a gross feedforward error attributable to a major process upset and results in corrective action by the faster acting deadband pH controller. To appreciate the logic of this combined control system it must be accepted that the stoichio-

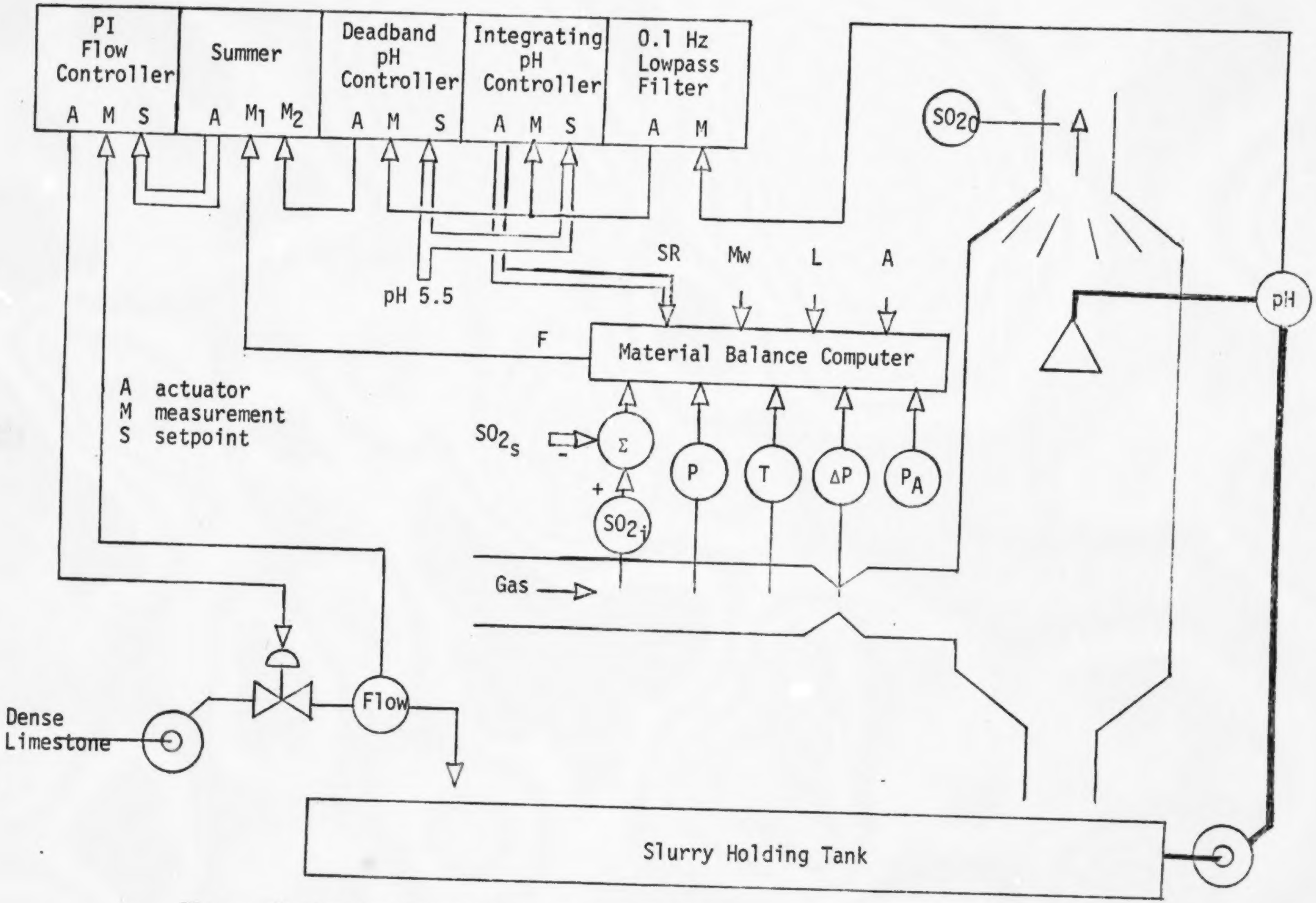

PI Flow Controller
A M S
Summer
A M_1 M_2
Deadband pH Controller
A M S
Integrating pH Controller
A M S
0.1 Hz Lowpass Filter
A M
SO_{2o}
pH 5.5
SR
Mw
L
A
F
Material Balance Computer
pH
A actuator
M measurement
S setpoint
SO_{2s}
−
+
Σ
SO_{2i}
P
T
ΔP
P_A
Gas
Dense Limestone
Flow
Slurry Holding Tank

Figure 7. Combined Stoichiometric And pH Limestone Addition Control System

metric ratio input as defined for material balance limestone feedrate control is actually the feedforward gain setting, that errors in feedforward stoichiometric control are primarily attributable to the slurry factors of Table 2 which pH measurements are sensitive to, and with the proper sense of direction (decreasing stoichiometric ratio with increasing pH) the integrating pH controller can adaptively trim this feedforward gain to track process anomalies and maintain accurate feedforward control.

Figure 8 illustrates this deadband pH and adpative stoichiometric ratio envelope control relationship. The solid gain curve for the deadband pH controller has a low value within the deadband zone, and an increasing value for external pH excursions. The stoichiometric ratio feedforward gain is adaptively adjusted along the vertical axis by the integrating pH controller, following a startup estimate which is preset for the stoichiometric ratio (SR) initial value. An accurate SR gain will coincide with the center of the pH deadband setpoint and zero pH gain.

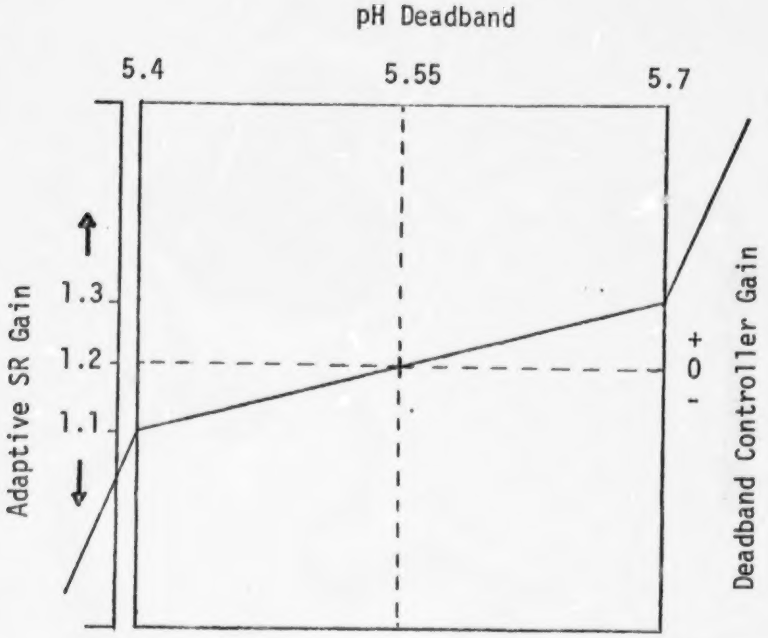


Figure 8. SR Vs. pH Envelope Control Relationship

FLUE GAS DESULFURIZATION SLUDGE DISPOSAL - THE SHAWNEE FIELD EVALUATION PROJECT

by

Paul R. Hurt, Paul P. Leo, Jerome Rossoff, and Jack R. Witz
The Aerospace Corporation
Los Angeles, California

Presented at the

U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory - RTP
Fifth Industry Briefing Conference on
Results of Lime/Limestone Wet Scrubbing Programs
Raleigh, North Carolina

December 5, 1979

ABSTRACT

This paper summarizes activities and results of the U.S. Environmental Protection Agency Flue Gas Desulfurization (FGD) field evaluation project at the Tennessee Valley Authority Shawnee Steam Plant near Paducah, Kentucky. This project, which has been in operation for the past 5 years, is evaluating various methods of FGD disposal by ponding and landfilling with particular interest on water quality and land reclamation. Ten small disposal sites are being evaluated for the disposal of lime and limestone sludges from nonregenerable scrubbers. Two ponds contain untreated sludge, three contain chemically treated sludge, and two contain untreated sludge with underdrainage; additionally, there are three gypsum stacks. Data are presented that show: (a) the ability of chemical treatment to produce a structural material, reduce concentrations of major species in leachate, reduce the coefficient of permeability, and shed water so that seepage can be prevented; (b) the value of underdraining untreated sludge in an impoundment to control seepage and to dewater the material so that it becomes a load-bearing structure; and (c) the detrimental effects of weathering, rewetting, and reslurrying on stacked gypsum filter cake. Further study is being made of gypsum to evaluate workable methods. Cost estimates are given for the disposal methods being evaluated at Shawnee.

SECTION I. INTRODUCTION

This paper summarizes activities and results of the EPA Shawnee Flue Gas Desulfurization (FGD) Field Disposal Evaluation Project that is in its sixth and final scheduled year of operation. This project is evaluating various ponding and landfill alternatives for the disposal of by-products from nonregenerable FGD systems. The effects of various disposal techniques, scrubbing operations, weather, soil interactions, and field operation procedures on the environmental quality of the disposal site are considered with the goal of determining environmentally sound disposal methods. Water quality and land reclamation are of principal interest and, to that end, periodic sampling, analyses, and assessments are being conducted of leachate, supernate, runoff, groundwater, and soil and sludge cores.

Currently, 10 field sites are being evaluated in this project which is located at the Tennessee Valley Authority (TVA) Shawnee Power Station near Paducah, Kentucky. Waste materials for this project are produced by each of two scrubber systems functioning in parallel at this station as an EPA/TVA test facility. The scrubbers [i.e., a turbulent contact absorber (TCA) and a venturi and spray tower (VST)] are each capable of treating up to 10 MWe of flue gas. Sludges from both scrubbers using lime and limestone as the SO_2 absorbent, and sludge oxidized to gypsum are being evaluated. This project provides a broad data base for the evaluation of flue gas SO_2 control by combining evaluations of scrubber operation, sludge disposal, and laboratory analyses.

Of the 10 sites in operation, 8 are small ponds ranging in size up to 0.10 acre with a sludge depth of 3 to 4 ft. These are excavated in a highly impervious clay soil that extends about 45 ft to the water table. Two ponds are filled with untreated sludge -- one from a lime and the other from a limestone scrubber, three contain chemically treated lime or limestone sludges, and three contain untreated sludge and are equipped with underdrainage. One of the latter contains a sludge that has been oxidized to gypsum, filled to capacity, then piled above grade. All sludges contain about 40 wt % fly ash except for the gypsum site. Two surface sites currently under evaluation are stacked with gypsum sludges -- one contains adipic acid.

The project planning, coordination, analysis, and reporting are performed by The Aerospace Corporation. Site construction, maintenance, coring, water sampling, and water analysis are performed by the TVA. The EPA Project Officer is Julian W. Jones, Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina.

SECTION II. PROJECT DESCRIPTION

Numerous reports and papers have been written on the Shawnee FGD Sludge Disposal Field Evaluation Project. Annual reports that provide detailed data defining site layout, sampling and analytical requirements, chemical and physical characteristics, disposal procedures, and cost estimates are given in References 1 through 6. Recent summaries of this project are given in References 7 through 9. This paper highlights some significant results and includes a brief statement of the findings.

Table 1 lists all project disposal sites, provides pertinent information, and gives the current site status.

The following sections separately discuss each type of disposal method being evaluated -- untreated ponding, chemical treatment, underdrainage, and forced-oxidation to gypsum. Cost estimates and overall findings are also provided.

Table 1. Shawnee Disposal Sites

Site	Fill Date	Scrubber Type^{a}	Sludge		Solids^{c} Content (wt %)	Treatment	Remarks
			Absorbent	Source^{b}			
A	10/8/74	VST	Lime	F	46	Untreated	Out of service 4/15/76
Al	5/10/76	VST	Lime	F	46	Untreated	Control pond, transferred from Pond A
B	4/15/75	TCA	Limestone	CU	38^{d}	Dravo	Underwater disposal
C	4/23/75	VST	Lime	CE	55^{d}	IUCS	Pond converted to runoff mode 3/79
D	2/5/75	TCA	Limestone	CU	38	Untreated	Control pond
E	12/7/74	TCA	Limestone	CU	38^{d}	Chemfix	Closed 11/77
F	2/3/77	TCA	Limestone	CU	47	Untreated	Underdrained pond, closed 11/77
G	10/5/76	VST	Lime	CE	47	Untreated	Underdrained pond
H^{e}	9/2/77	VST	Limestone	CU	33	Untreated	Pond
	9/30/77	VST	Limestone	F	86	Untreated	Surface site, unreacted limestone 13% dry wt
J^{e}	12/31/78	VST	Limestone	F	81	Untreated	Surface site, has adipic acid
K^{e}	3/29/79	VST	Limestone	F	80	Untreated	Surface site, unreacted limestone 25% dry wt

[a] Venturi and spray tower (VST); turbulent contact absorber (TCA)

[b] Filter (F), clarifier underflow (CU), and centrifuge (CE)

[c] Pond H is ash-free. All others: fly ash is approximately 40 wt % of solids content

[d] Prior to chemical treatment

[e] Forced-oxidized to gypsum

SECTION III. UNTREATED SLUDGE PONDING

Two control ponds for lime and limestone sludge disposal were installed and are being monitored principally for the determination of chemical characteristics of leachate. These ponds, which are identified in Table 1 as A/A1 and D, are totally saturated (except for periods of extreme drought at Shawnee). After 5 years, they remain nonstructural and continue to seep water to the subsoil. A current photograph of Pond D is shown in Figure 1.

The initial leachate of each pond contained a total dissolved solids (TDS) concentration equal to that of the input liquor. After approximately 1 year the chloride content of the sludge had been essentially depleted, and, thereafter, the leachate was saturated in gypsum (Figure 2) and has remained so after 5 years. Analyses have been made for the concentration of the more significant minor species present in these sludges (i.e., arsenic, mercury, lead, selenium, and boron). Except for boron, which decreases steadily with time, (Figure 3) the minor species have shown only slight reductions after 5 years of seepage.

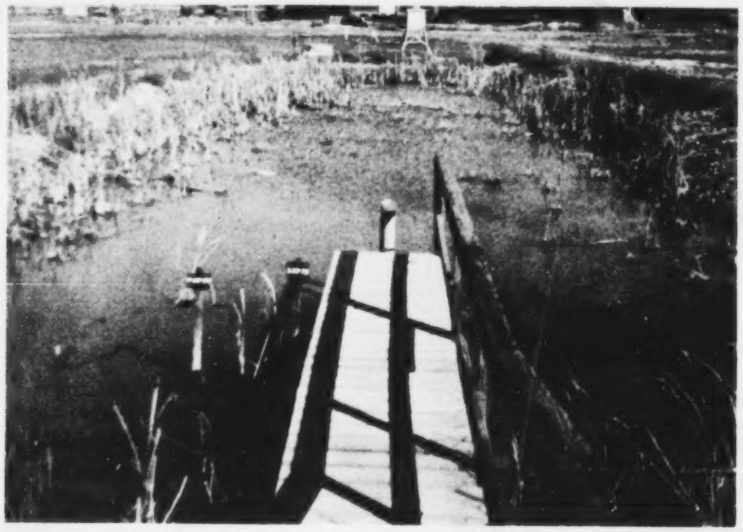


Figure 1. Pond D -- Untreated Limestone Sludge, Current Condition

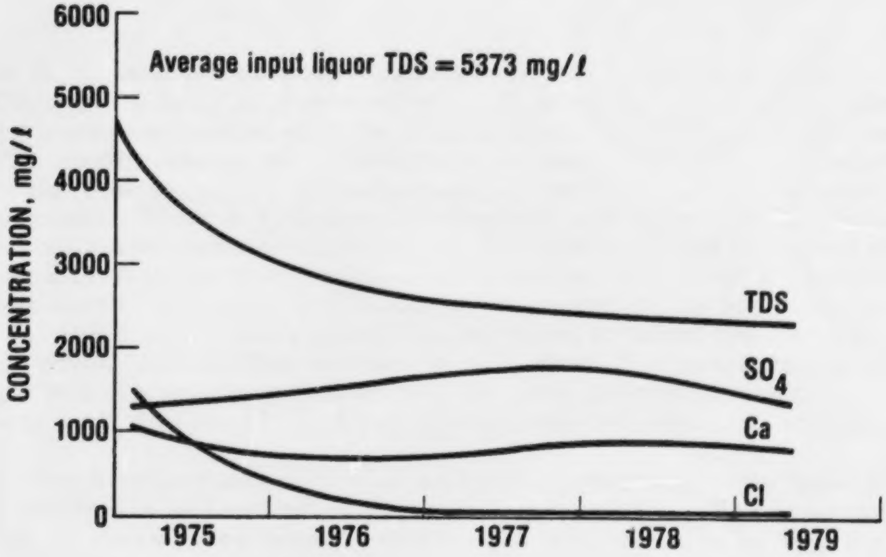


Figure 2. Pond D Leachate -- Concentration of TDS and Major Species

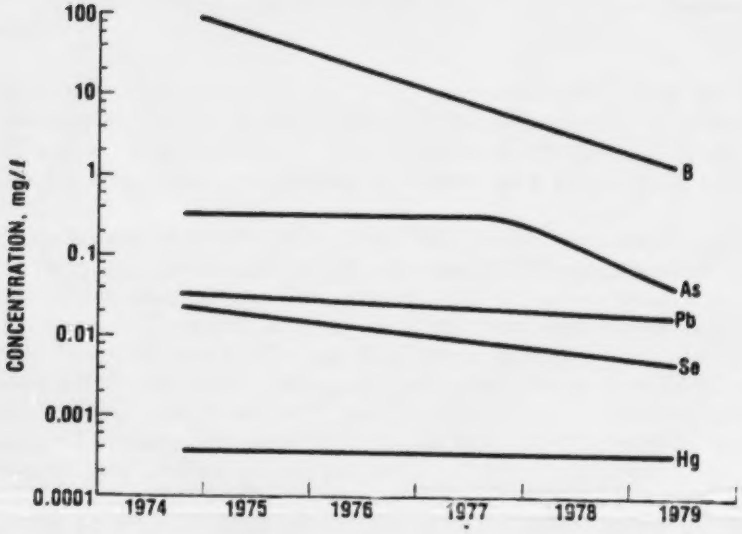


Figure 3. Pond D Leachate -- Concentration of Minor Species

SECTION IV. CHEMICALLY TREATED SLUDGES

Ponds B, C, and E contain sludges which were chemically treated by Dravo, IUCS., and Chemfix, respectively. In general, all of these chemically treated processes converted the sludge to a structually sound material, decreased the coefficient of permeability by at least one order of magnitude, and reduced the concentration of major species in the leachate by approximately 50 percent. Pond B simulates a disposal site in which the sludge cures underwater and remains underwater thereafter, Pond C represents a "low spot" in a landfill in which rainwater collects, and Pond E represents a landfill that traps rainwater that collects in a sump at its lower end. Figures 4, 5, and 6 show these ponds during their original placement, at an intermediate point, and as they appear today. Pond B remains as originally configured, Pond C was converted during the past year to a runoff configuration, and, in late 1977, Pond E was covered with clay and planted.

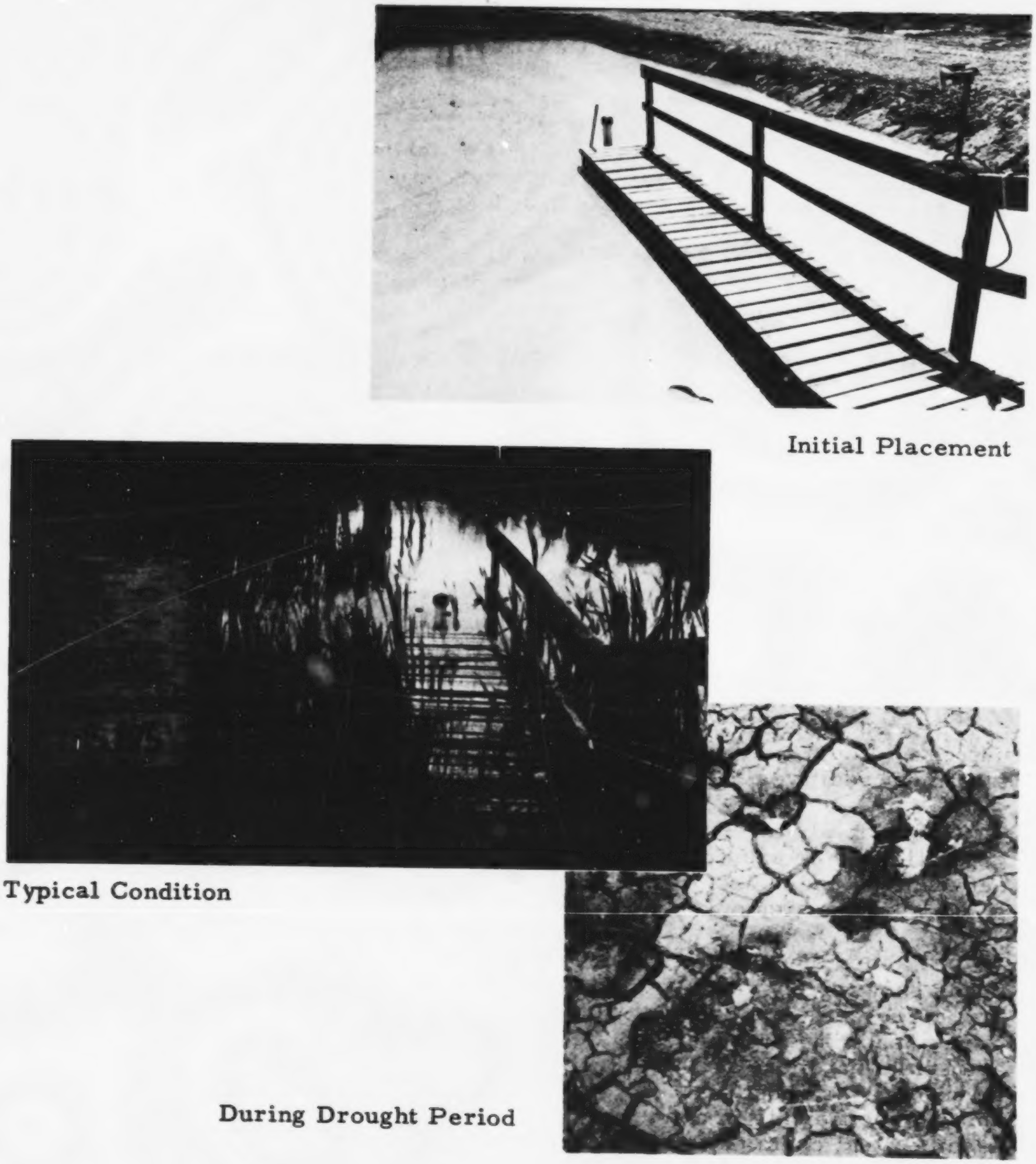
Plots of concentrations of major and minor species in the leachate for these three ponds are presented in Figures 7 through 12. The maximum TDS concentration in these three ponds, which occurred initially or shortly thereafter, was approximately one-half the TDS concentration in the input liquor.

The minor species concentrations are shown in Figures 8, 10, and 12 and are summarized in Table 2. Reductions of concentration with time are shown. In some instances, the concentration of constituents in the leachate of a treated sludge is somewhat higher than the concentration in the input, untreated liquor. This probably results from trace elements present in the treatment additives.

Physically these materials (Table 3) all demonstrate high bearing capacities. For example, Pond B exhibits ultimate bearing capacities between 150 and 300 psi, and Ponds C and E exceed 300 psi. Other pertinent physical properties of chemically treated sludges are given in Table 4.

Laboratory analyses of materials from these three chemically treated ponds show widely varying coefficients of permeability as a result of cracks that appear in some core samples. In general, it appears that the coefficient of permeability for Ponds C and E are approximately the same and that they are somewhat more "impermeable" than the Pond B material. Although all three register about 10^{-5} cm/sec, several samples from Ponds C and E showed coefficients in the range of 10^{-6} to 10^{-7} cm/sec. Therefore, it is assumed that chemical treatment will decrease the coefficient of permeability of a sludge by at least one order of magnitude.

Pond E (Figure 6) was closed after 3 years of evaluation by evacuating the leachate well and the sump at the low end, covering the pond with clay, and vegetating it. Procedures used were similar to those described for Pond F in Section V.



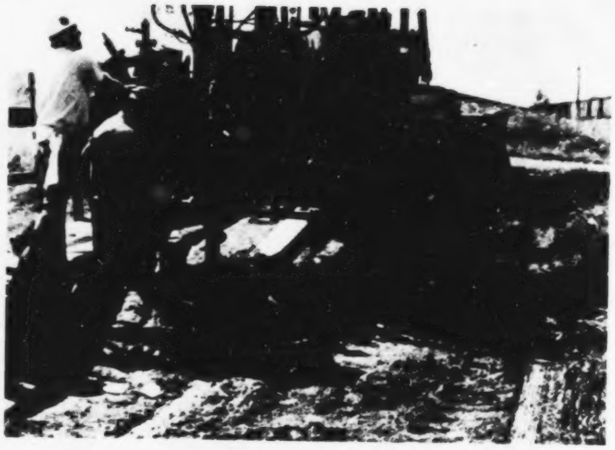
Initial Placement

Typical Condition

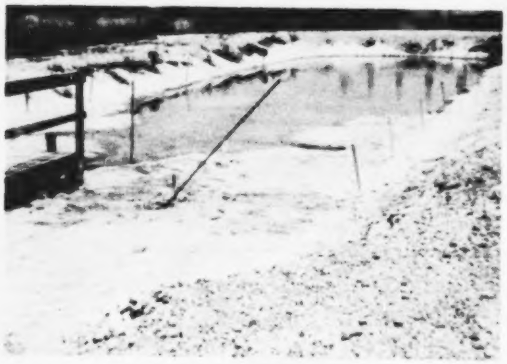
During Drought Period

Figure 4. Pond B

Supporting Coring Rig



Containing Trapped Water



After Conversion to Runoff Configuration

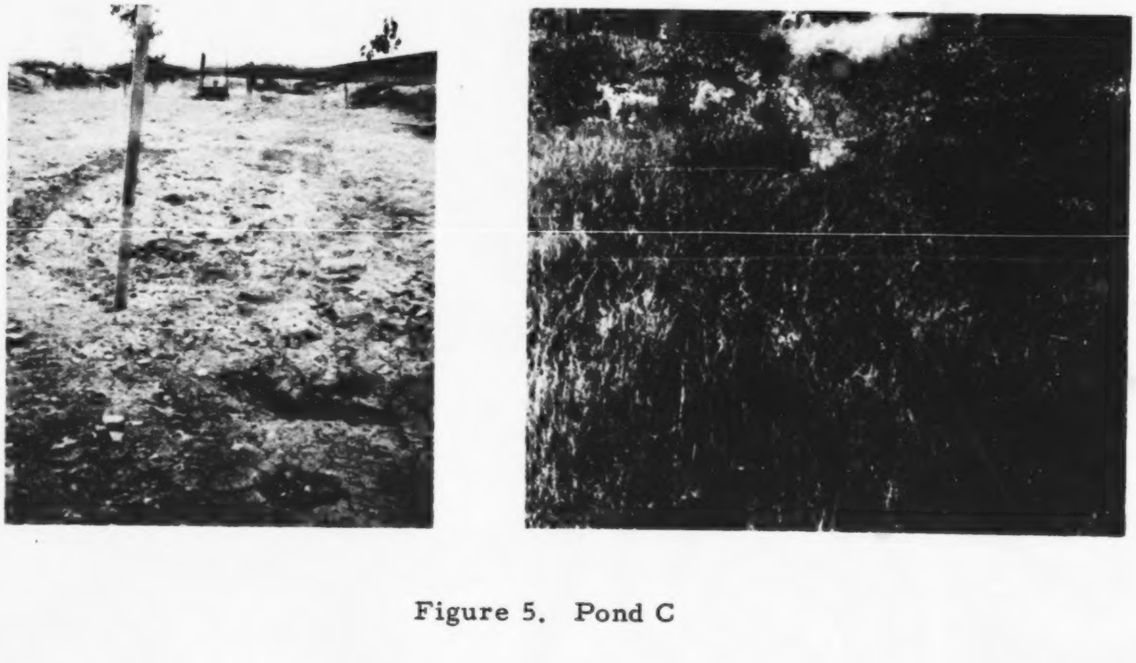
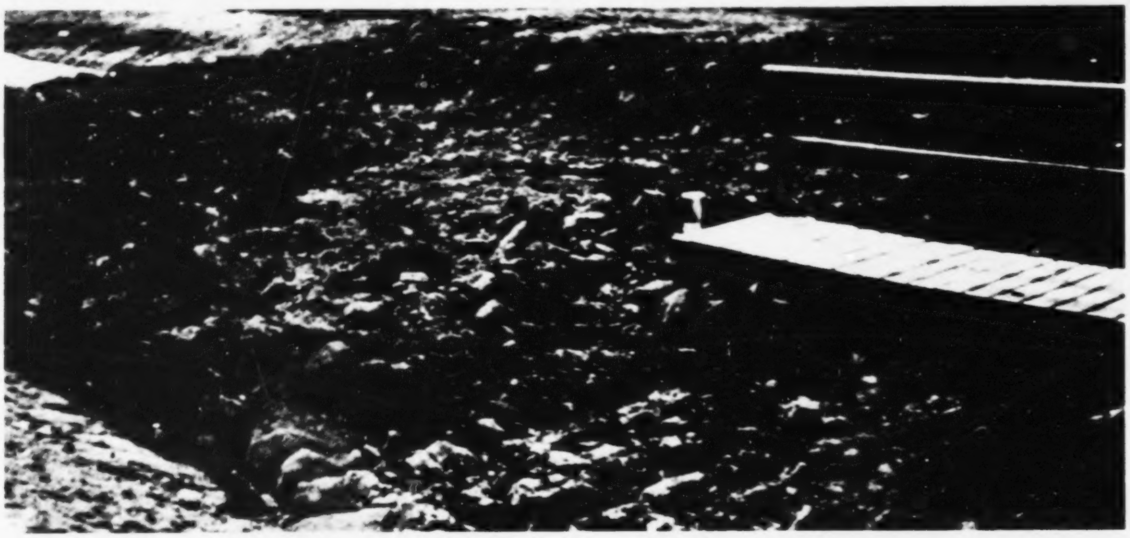
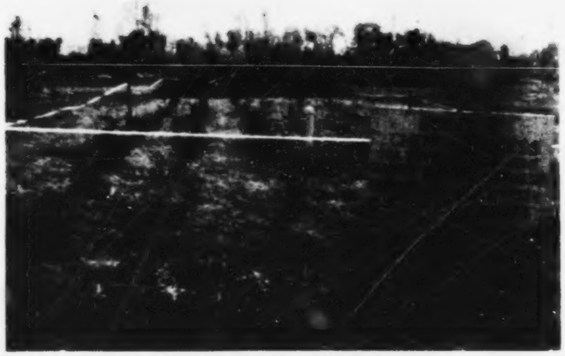


Figure 5. Pond C



Initial Configuration

Containing Trapped Water and Supporting Drill Rig



Current Closed Configuration

Figure 6. Pond E

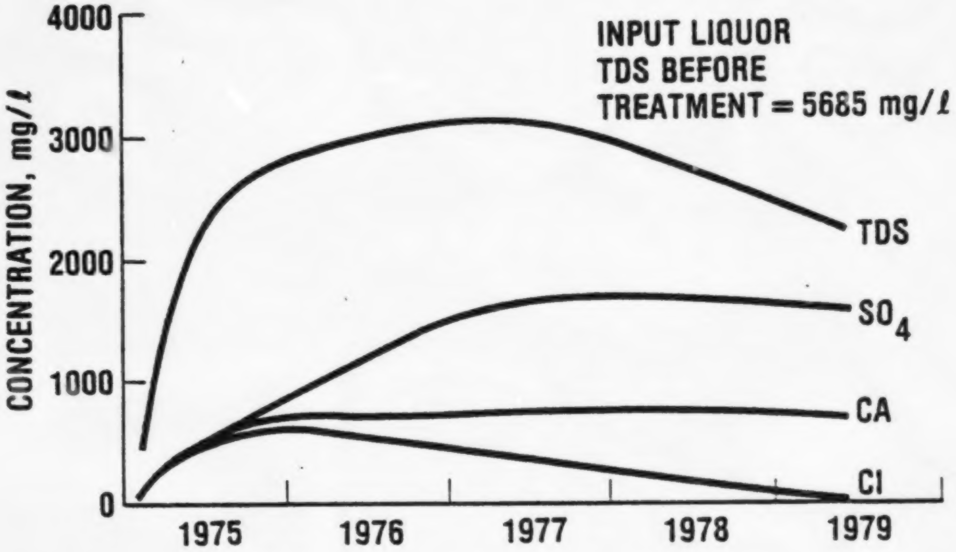


Figure 7. Pond B Leachate -- Concentration of TDS and Major Species

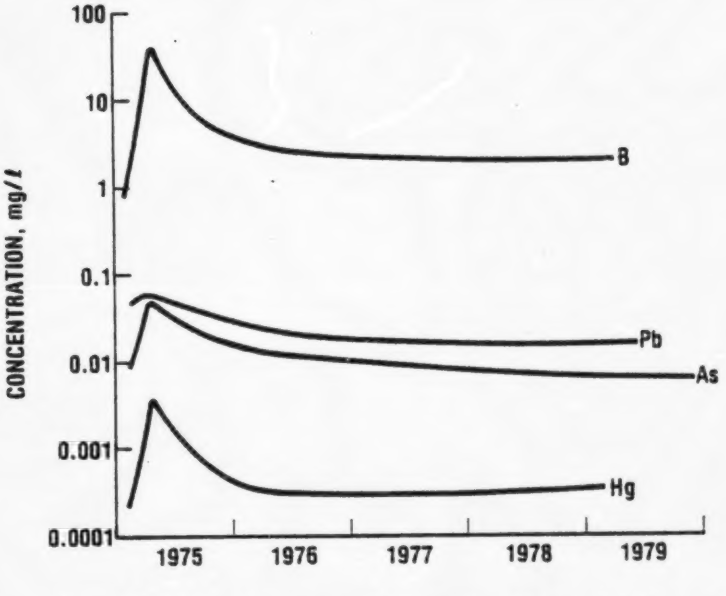


Figure 8. Pond B Leachate -- Concentration of Minor Species

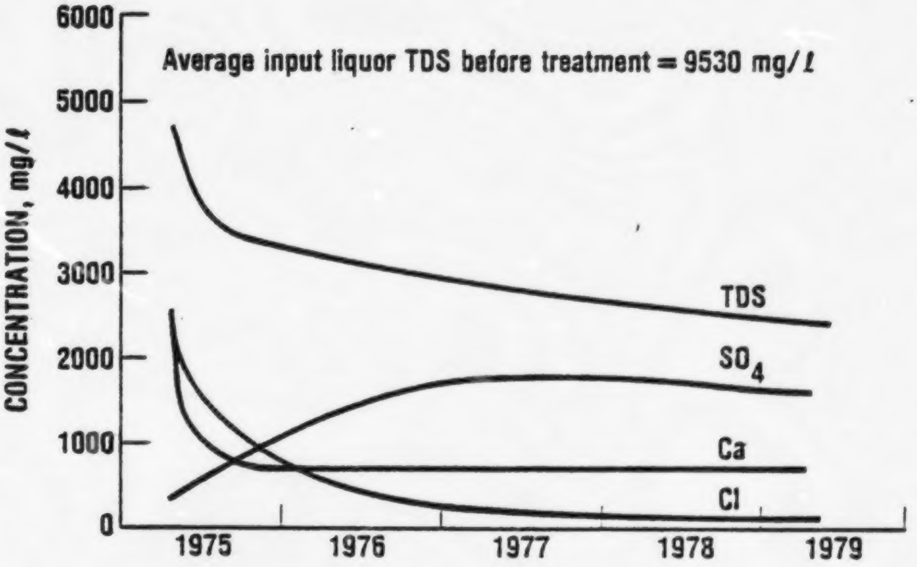


Figure 9. Pond C Leachate -- Concentration of TDS and Major Species

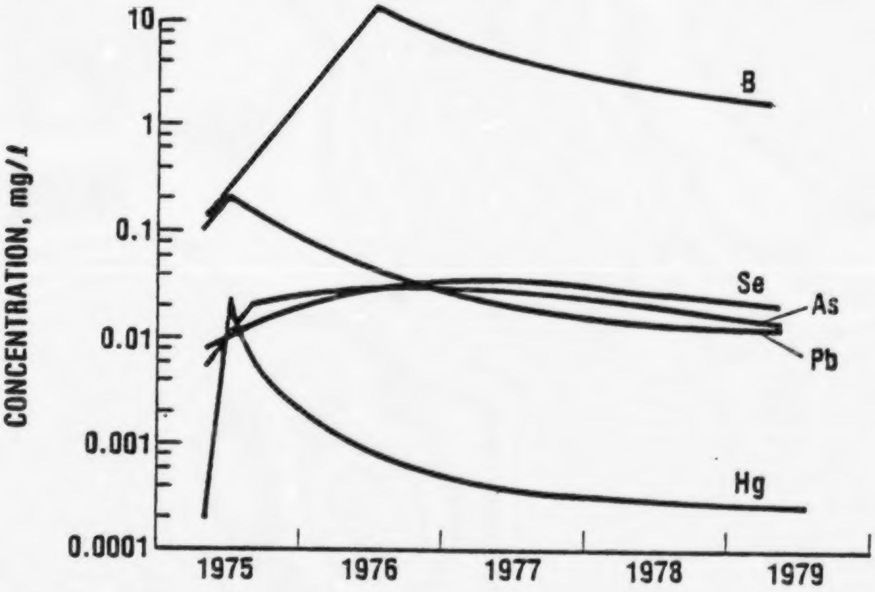


Figure 10. Pond C Leachate -- Concentration of Minor Species

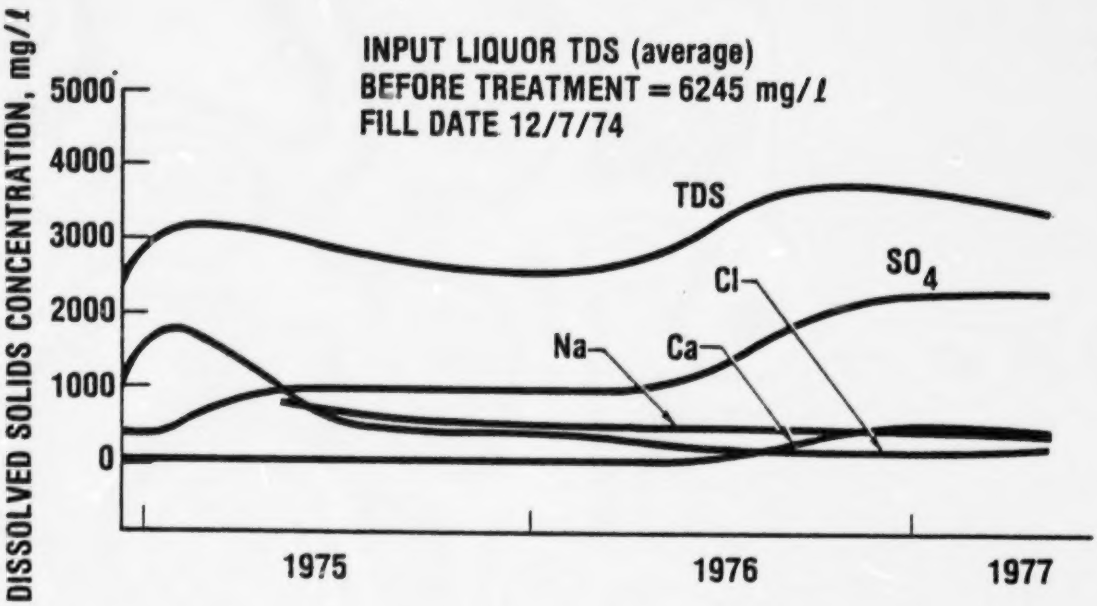


Figure 11. Pond E Leachate -- Concentration of TDS and Major Species

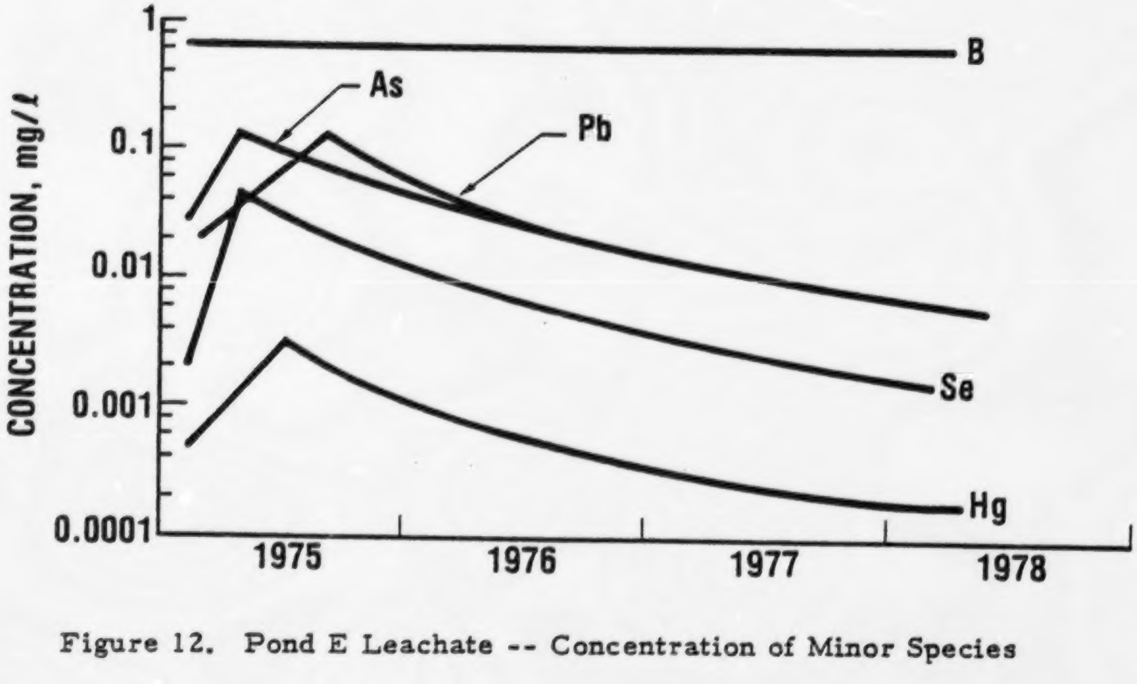


Figure 12. Pond E Leachate -- Concentration of Minor Species

Table 2. Sludge Leachate -- Summary of Concentrations of Minor Species

Element	Untreated						Chemically Treated								
	Pond A			Pond D			Pond B			Pond C			Pond E		
	L_i	L_o	L_t	L_i	L_o	L_t	L_i	L_o	L_t	L_i	L_o	L_t	L_i	L_o	L_t
As	0.024	0.08	0.002	0.24	0.11	0.4	0.004	0.04	0.007	0.002	0.035	0.02	0.003	0.15	0.006
B	44	48	5	90	89	1.3	97	50	2	34	15	2	80	0.7	0.7
Pb	0.49	0.05	0.01	0.26	0.03	0.02	<0.02	0.08	0.015	<0.01	0.2	0.015	<0.01	0.15	0.01
Hg	<0.0001	0.001	0.0002	<0.0001	0.003	0.0002	0.00024	0.004	0.003	<0.00008	0.025	0.003	0.00033	0.003	0.0002
Se	0.005	0.004	0.002	0.040	0.10	0.003	0.02	0.2	0.005	0.018	0.03	0.02	0.014	0.05	0.0015

L_i = Untreated input

L_o = Treated, initial leachate

L_t = Treated, most recent leachate

Table 3. Ultimate Bearing Capacity of Ponded Sludge

Pond	Treatment	Ultimate Bearing Capacity, psi
B	Dravo	150 to 300
C	IUCS	>330
E	Chemfix	300 to >330
F[a]	Untreated	60 to 75
G[a]	Untreated	180 to 240
H[b]	Untreated	135 to >330

[a] Underdrained

[b] Clarifier underflow, underdrained

Note: Shawnee clay soil has an ultimate bearing capacity of from 240 to 300 psi.

Table 4. Physical Characteristics of Impounded Treated Sludge Cores

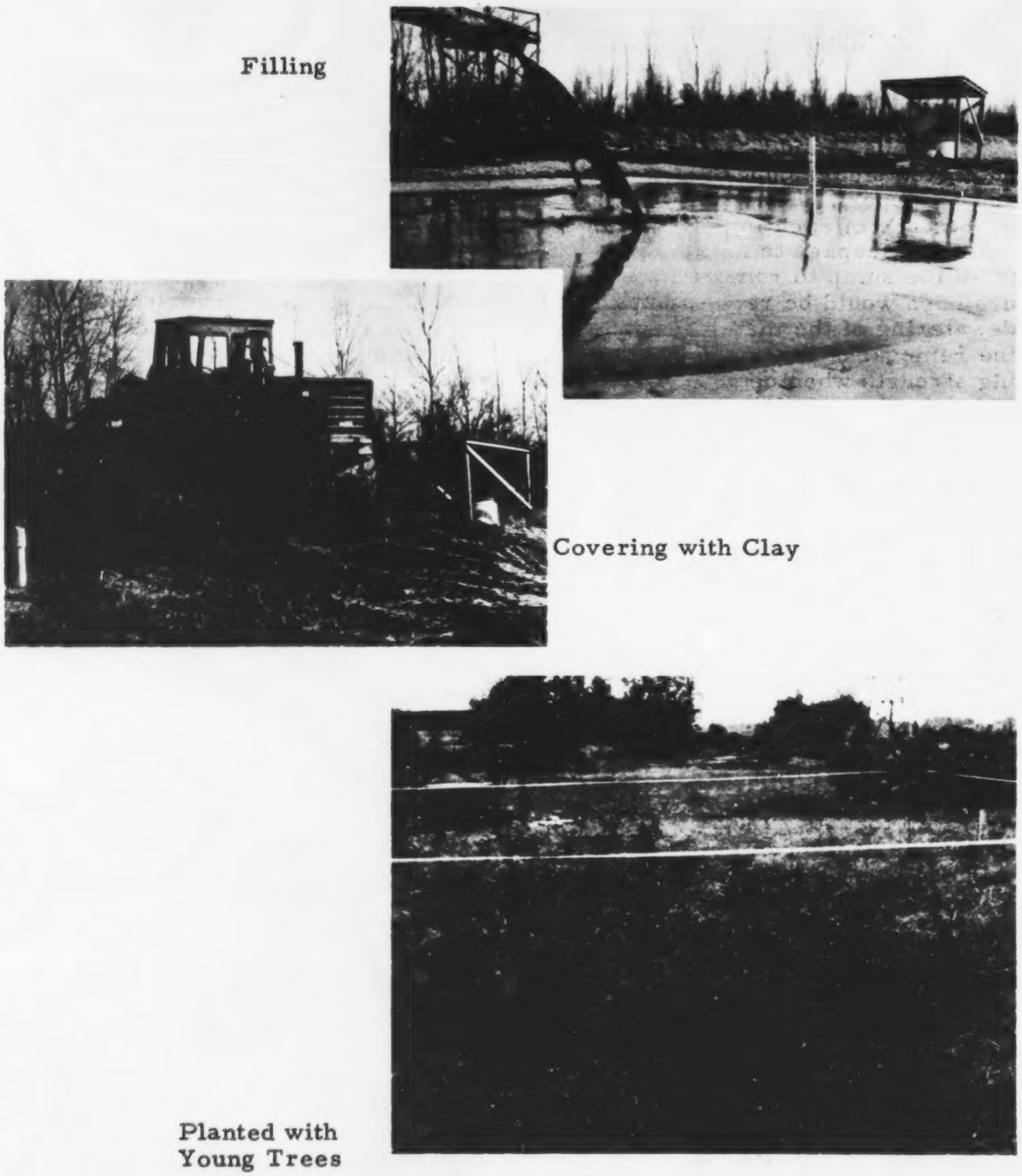
Pond	Solids Content, wt %	Density (wet), g/cm^3	Volumetric Void Fraction
B	44	1.37	0.75
C	61	1.52	0.63
E	50	1.36	0.71

SECTION V. UNDERDRAINAGE OF UNTREATED SLUDGES

Ponds F and G and the initial phase of Pond H are evaluations of untreated sludge deposited in an underdrained impoundment (Figures 13, 14, and 15-upper). The underdrainage system consists of a 1-ft-thick sand layer at the base of each impoundment, under which are perforated plastic pipes that feed seepage to a sump through a gravity drain system. Water is pumped from the sump to remove all seepage (in an operational site the underdrainage would be recycled to the scrubber). The underdrainage enhances dewatering of the material such that the additional settling that results from the removal of the occluded water results in a material with very high bearing strength when contained in an impoundment. For example, as shown in Table 3, Pond F (limestone sludge) has an ultimate bearing capacity of 60 to 75 psi, Pond G (lime sludge) has 180 to 240 psi, and Pond H (gypsum clarifier underflow) has 135 to 330 psi. These ponds drain rapidly. They could support personnel during the pouring operation, and Ponds G and H supported wheeled vehicles within a day after placement. Pond F, the weaker of the three, did not drain as rapidly as the others but still attained high bearing strength. Also, after the gypsum clarifier underflow in Pond H had settled and physical analyses were made, the underdrainage was turned off and gypsum filter cake was stacked on the site (Figure 15) to simulate filter cake piled above ground without drainage. At Pond H the underdrainage is turned on every 2 months to obtain a 2-liter sample of leachate for chemical analysis.

After Pond F had been in service for 9 months, all water was removed from the underdrain sump and the site was covered with local clay, using a rubber-tired earth mover. This clay cap was contoured and planted with grass so that its surface would shed rain water. The cap was approximately 3-ft-thick at the center line and 2-ft-thick at the edges. In the spring of 1979 this cap was cut back by about 1 ft and was planted with young trees by the TVA for observation within a separate project.

During subsequent monitoring of this site several coring holes were left unfilled, and, as a result, water was directed to the base of the pond through those holes. After the water was removed from the holes and the underdrainage was evacuated, the holes were filled with clay and tamped. The pond bottom has remained dry. Thus, it is apparent that any closed sludge disposal site must be monitored to prevent rupture of the cap if rainwater is to be prevented from seeping through the sludge.

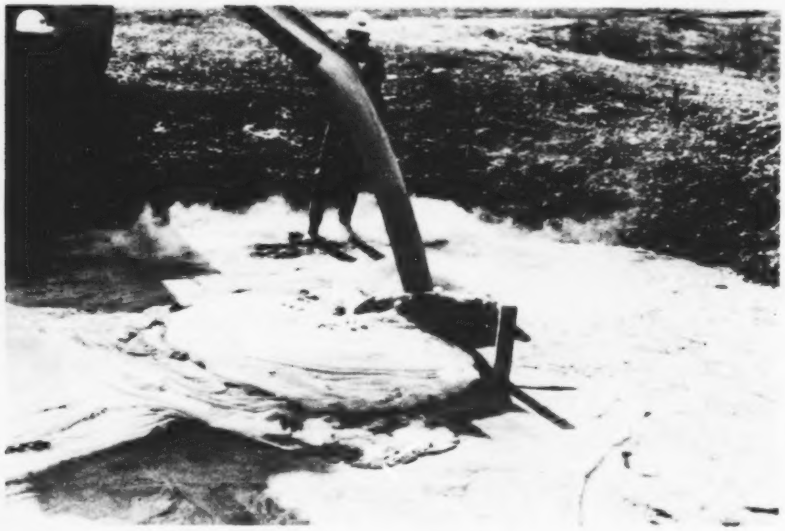


Filling

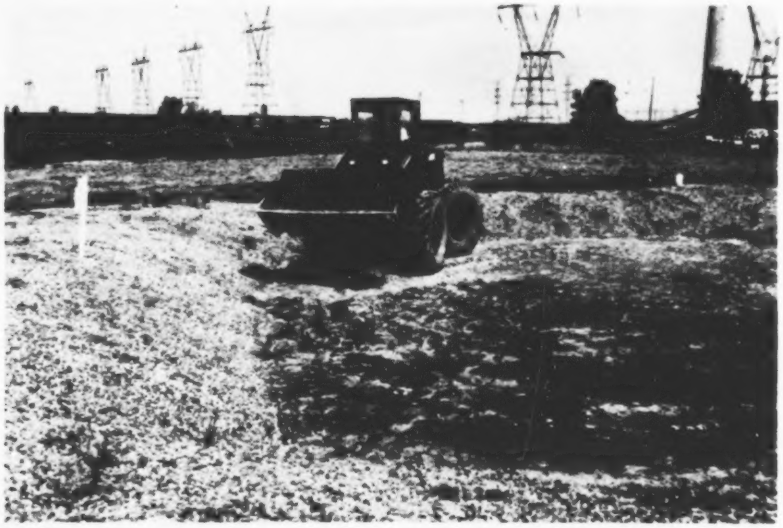
Covering with Clay

Planted with Young Trees

Figure 13. Pond F -- Untreated Limestone Sludge with Underdrain



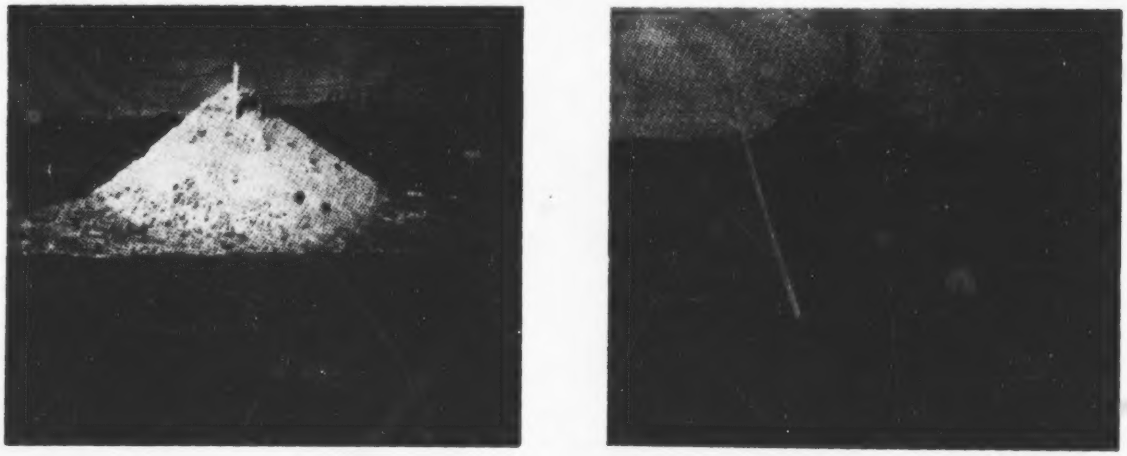
Filling



One Day After Filling

Figure 14. Pond G -- Untreated Lime Sludge with Underdrainage

One Day After Filling



After Placement

Now

Gypsum Filter Cake

Figure 15. Site H -- Gypsum, Clarifer Underflow Underdrained

SECTION VI. GYPSUM FILTER CAKE

Gypsum filter cake was stacked at Sites H, J, and K. Significant sludge characteristics are given in Tables 5 and 6. The filter cake at Site H contains 13 percent calcium carbonate and no fly ash; the material at Site J, in addition to having adipic acid additive, contains very little calcium carbonate and 35 percent fly ash; and at Site K the cake has 25 percent calcium carbonate and 28 percent fly ash. The cake at Site H is not considered typical because of its high percentage of unreacted limestone, but it was used because it was available at the time the site was initiated. The cake at Site J is more representative of an operational material. Finally, the cake at Site K, although highly irregular in that it has 25 percent unreacted limestone, is considered representative of the gypsum that will be produced at the TVA Widows Creek Power Station. These tests determined the impact of merely piling gypsum on the ground. Observations have been made as to slump, runoff, water quality, surface crust formation, crust bearing strength, and attendant strength loss when moisture content is increased. Views of these three sites are presented in Figures 15, 16, and 17.

At Sites H and J the filter cake was stacked so that a natural slope occurred on a conical shape producing a surface slope of about 35 deg to the horizontal. A crust caused by rehydration of the gypsum filter cake formed on the surface of these piles. The crust hardened and shed water except in areas where shrinkage cracks occurred. Field penetrometer readings were taken of ultimate bearing capacity on these two sites. At this time, crust thickness varies from approximately 2 in. at the peak of the cone to 5 in. at mid-height, to 11 in. at the base. Crust strength on both sites increases from the peak to the base (Table 7). However, immediately below the crust (which has a solids content of 81 to 85 percent) the gypsum filter cake has absorbed rainwater which has reduced the solids content to about 65 percent and the bearing capacity to 0 to 30 psi. At this time, these two sites have become hard crusted cones supported by a weak, moist core.

Chemical analyses have been made of underdrainage and runoff samples from Site H. The underdrainage is a typical untreated sludge leachate as expected (Figure 18). The runoff has a TDS slightly in excess of 2000 mg/ℓ and a total suspended solids that ranges between 4 and 300 mg/ℓ, indicating that runoff would have to be controlled so that it does not seep into ground water supplies or become discharged into streams. It appears that slumping would be a problem with disposal of gypsum in this manner. Within the first 18 months of observation, approximately 20 percent of the mass flowed to the base of the site (Figures 15 and 16). The highly errosive nature of this crust is evident in the close-up picture (Figure 16) taken of Site J. Several tests were made with a D-8 Caterpillar at Site J to determine the maximum angle at which the vehicle could negotiate the slope of this material. During the test, the cleats of the tractor broke through

Table 5. Gypsum Disposal Sites Input Data -- Limestone Absorbent All Sites

Site	Placement Dates	Quantity, cu yd	Underdrain	Ash[a]	Adipic Acid
H[b] (lower)	8/19/77 to 9/2/77	84	Yes[c]	<1	No
H[d] (upper)	9/19/77 to 9/28/77	122	No	<1	No
J[d]	10/10/78 to 12/31/78	500	No	40	Yes
K[d]	3/3/79 to 3/29/79	150	No	31	No

[a] Wt percent solids before oxidation

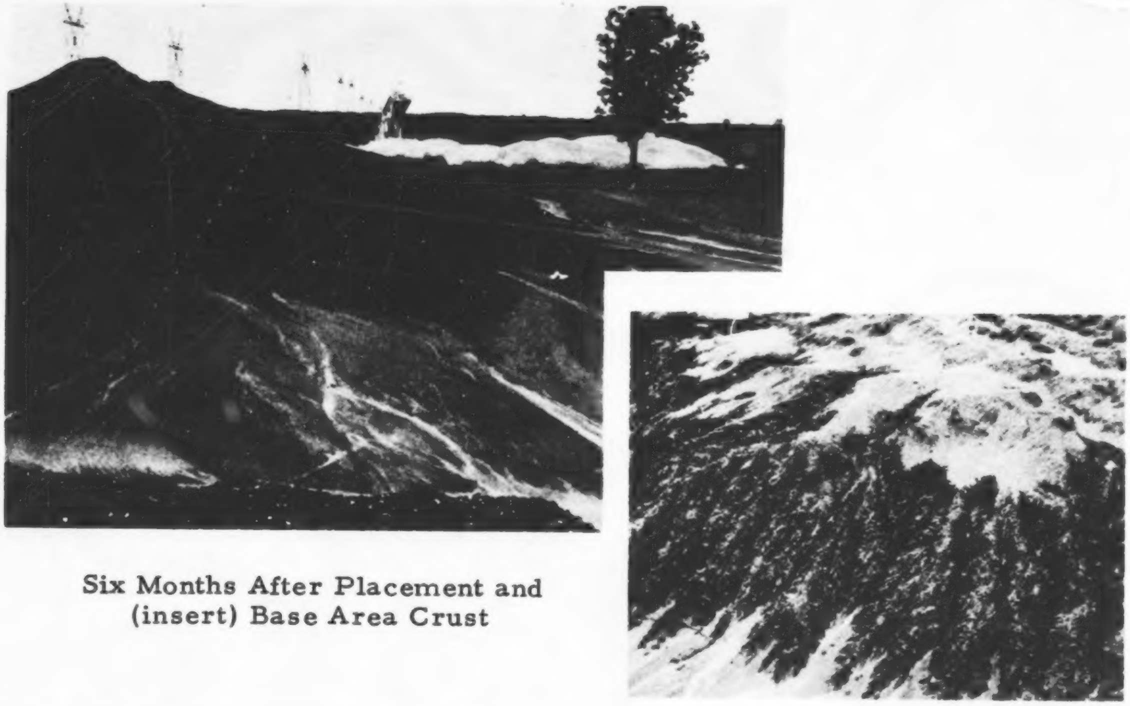
[b] Clarifier underflow

[c] Underdrain closed on 10/12/77. Reopened approximately every 2 months to take 2-liter sample.

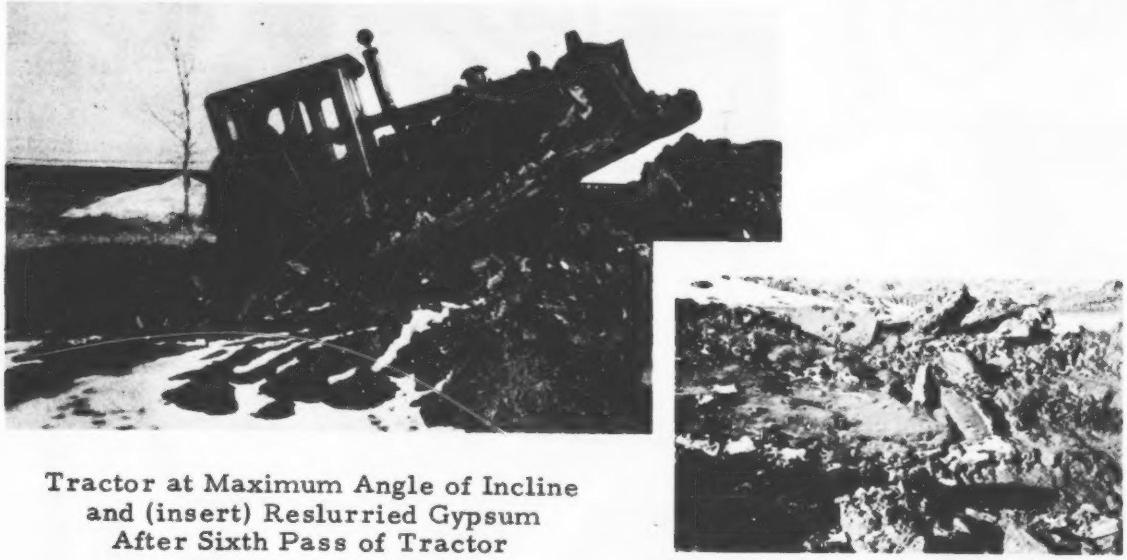
[d] Filter cake

Table 6. Solids Analysis -- Gypsum Disposal Sites

Site	Content, wt % solids				
	Input Sludge	Calcium			Fly Ash
		Sulfite	Sulfate	Carbonate	
H Clarifier Underflow	33.0	0.2	98.0	1.8	--
H Filter Cake	86.0	0.1	86.8	13.1	--
J Filter Cake	80.7	0.1	63.7	1.2	35.0
K Filter Cake	79.9	0.1	47.1	25.0	27.8



Six Months After Placement and (insert) Base Area Crust



Tractor at Maximum Angle of Incline and (insert) Reslurried Gypsum After Sixth Pass of Tractor

Figure 16. Site J -- Gypsum Filter Cake

Figure 17. Site K -- Gypsum Filter Cake

Table 7. Ultimate Bearing Capacity of Gypsum Filter Cake Crust

Limestone absorbent; forced-oxidized to gypsum.

Crust thickness varied from 2 in. at peak to 5 in. at mid-height to 11 in. at base. Crust contains 81 to 85 percent solids. Below crust, percent solids is approximately 65 percent with a bearing capacity of 0 to 30 psi.

Site	Ultimate Bearing Capacity at Pile Locations, psi		
	Peak	Mid-height	Base
H (13% unreacted limestone)	60	60 to 300	240 to 450
J (adipic acid added)	90 to 330	150 to 420	450

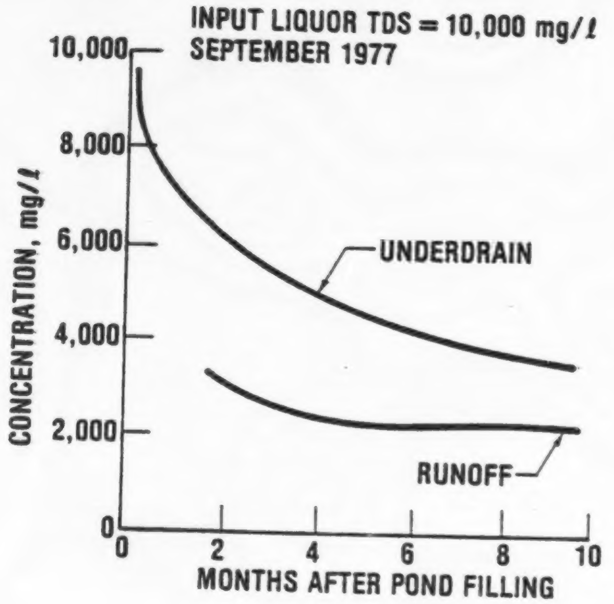


Figure 18. Pond H -- Concentration of TDS in Underdrain and Runoff

the crust, and at an angle of 17 deg to the horizontal it lost traction. After approximately six passes over the same spot, the material became highly liquid and unable to support personnel (Figure 16).

The material on Site K has been dumped in piles with no specific shape to be formed. Tests will be made to determine the benefits, if any, of compacting this material to enhance its structural characteristics.

Load bearing strengths of untreated sludges, including gypsum, as a function of moisture, fly ash content, and sludge origin (power plant, type of absorbent) are presented in Figure 19 for numerous untreated sludges. Among other considerations, the data highlight the criticality of solids content on the load bearing strength of untreated sludges. This is of particular importance when considering the disposal of slurried gypsum, gypsum filter cake, or untreated sludges that are underdrained because these types of disposal depend on dewatering to attain a desired material bearing strength. Figure 19 shows that there is a very short range of solids content for any given sludge over which its strength will decrease rapidly from an extremely high value to zero. When there is an external application of water to a disposal site (e.g., rainfall) there should be operating or maintenance provisions to prevent absorption of the water by the sludge or to remove the absorbed

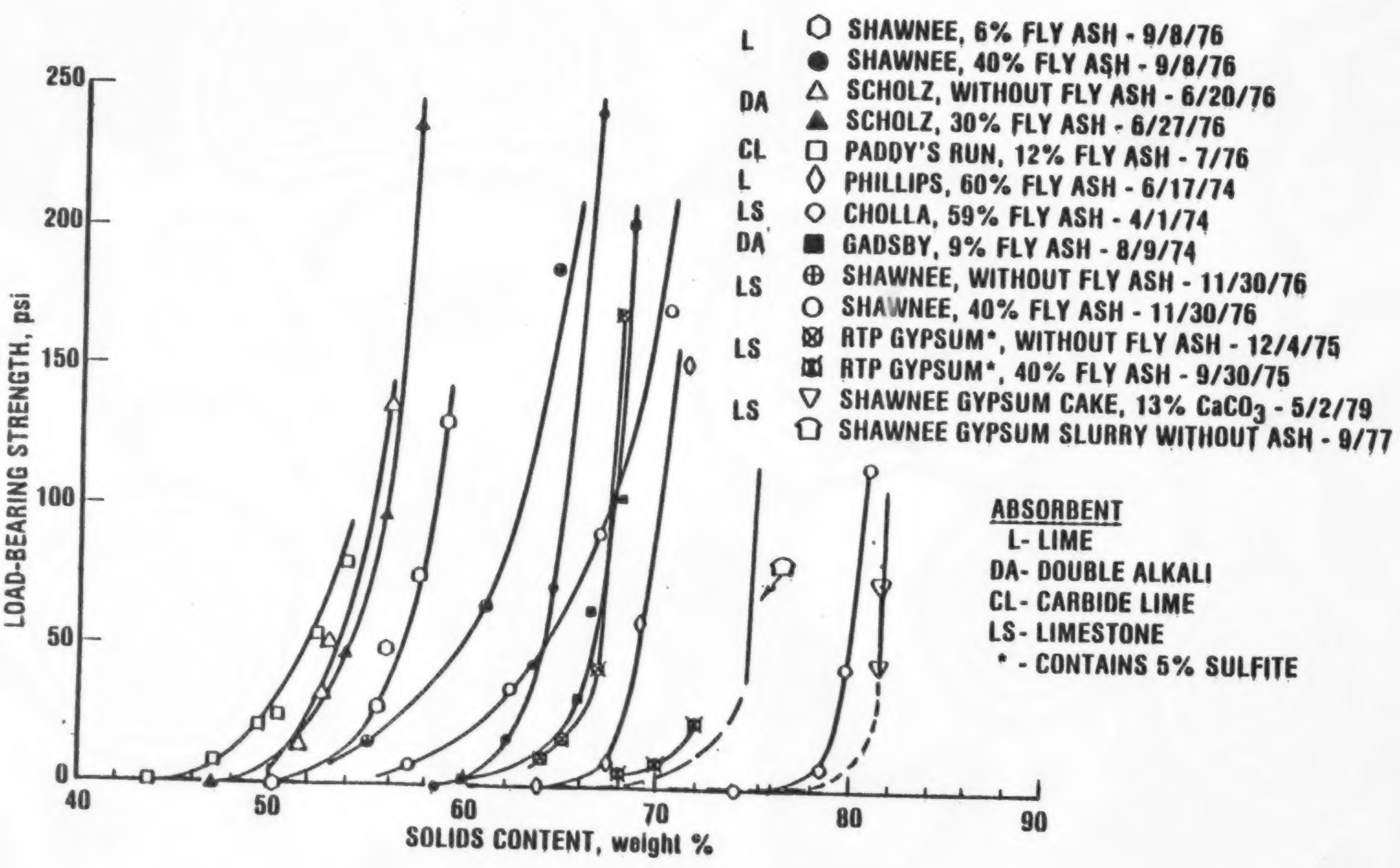


Figure 19. Load-Bearing Strength as a Function of Moisture, Fly Ash Content (wt % dry), and Sludge Origin

water if it should occur. Finally, the critical value for gypsum is not a single solids content for all sludges. It appears that this value ranges between the low 70's and the low 80's, with this variation being dependent upon ash content, unreacted limestone content, degree of oxidation, and original method of dewatering (i.e., filtering or settling/drainage). Crystal size and shape also could affect this value.

SECTION VII. COST ESTIMATES

Periodic estimates are made of projected cost for sludge disposal by various methods related to the type of disposal evaluation being conducted at Shawnee. Cost estimates for four different disposal methods are given in Table 8.

Table 8. Cost Estimates

All sludge includes fly ash; estimates in mid-1980 dollars.

Disposal Method	Cost Estimate		
	mills/kWh	$/ton sludge (dry)	$/ton coal
Ponding - Untreated			
Indigenous Liner	0.65	5.90	1.75
Liner Added	1.40	12.50	3.80
Reclaimable Impoundment -- Untreated Sludge with Underdrainage	1.05	9.40	2.85
Ponding -- Slurried Gypsum			
Indigenous Liner	1.20	10.20	3.15
Liner Added	1.75	15.05	4.65
Chemically Treated -- Landfill	1.25	11.85	3.60

These costs are not the result of conceptualized designs, but are intended to be typical for the type of disposal indicated. The estimates given in Table 8 are in mid-1980 dollars and are updates of values presented previously, except for the underdrain impoundment cost which is new. The underdrainage method indicates an approach not using oxidation equipment, clarifiers, or filters. The major underdrainage system cost is the acquisition and placement of a porous bed consisting of sand, gravel, or bottom ash, or the equivalent. The underdrained piping system consists of perforated plastic pipes placed approximately 130 ft apart. Because of the large spacing between the pipes, their cost is a small percentage of the total system cost.

SECTION VIII. FINDINGS

1. Pond B, containing a Dravo chemically treated sludge which can be disposed of under water, will support construction vehicles. Site C, containing IUCS chemically treated sludge, is structurally sound and has demonstrated the ability to shed water and prevent seepage. Pond E, containing Chemfix chemically treated sludge, is structurally sound. It was closed (see Finding 7); however, if it had been reconfigured to a runoff condition as Pond C was, it should have also shed water and prevented seepage, based on comparable coefficients of permeability.

2. Chemical treatment improves the coefficient of permeability by at least one order of magnitude, and reduces concentrations of major species in leachate by about 50 percent. Concentrations of trace elements being monitored in the leachate of chemically treated sludges show positive and negative variations from concentrations in the input liquor. Concentrations greater than input conditions may result from additional trace elements present in the chemicals added in the treatment process (e.g., cement, fly ash, lime).

3. The concentrations of TDS in the leachates of the chemically treated and untreated ponded wastes have approached levels which approximate saturation with gypsum.

4. The groundwaters associated with all ponds show no effect attributable to chemically treated and untreated waste disposal.

5. Runoff from sulfite sludge which has been force-oxidized to gypsum shows a TDS concentration between 2500 and 3200 mg/ℓ and the total suspended solids in the same runoff varies between 4 and 300 mg/ℓ. To maintain structural strength in the gypsum material, the site should be managed to shed water so that it is not allowed to reslurry by wetting.

6. The underdraining of untreated sludge (Ponds F and G) results in a structurally sound material in an impoundment; all seepage and rainfall on these sites is controlled and operationally would be returned to the scrubber and reused.

7. Pond closure by draining and covering and contouring with clay material has been demonstrated to be sufficient to retire disposal sites (chemically treated Pond E and untreated/underdrained Pond F) so that they will support construction vehicles and shed water, thereby preventing seepage.

8. In general, cost estimates in mid-1980 dollars for disposal range from 0.65 to 1.25 mills/kWh [$5.90 to $11.85/ton of sludge (dry)]. Ponding on indigenous liners has the least cost and chemical treatment the highest. An unusual case such as ponding gypsum on a site at which a liner would have to be added would be even more expensive. A reclaimable impoundment containing untreated sludge with underdrainage would cost approximately 1.05 mills/kWh [$9.40/ton of sludge (dry)].

REFERENCES

1. J. Rossoff, and R.C. Rossi, Disposal of By-Products from Nonregenerable Flue Gas Desulfurization Systems: Initial Report, EPA-650/2-74-037a, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, May 1974.
2. R.B. Fling et al., Disposal of Flue Gas Cleaning Wastes: EPA Shawnee Field Evaluation: Initial Report, EPA-600/2-76-070, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, March 1976.
3. J. Rossoff et al., Disposal of By-Products from Nonregenerable Flue Gas Desulfurization Systems: Second Progress Report, EPA-600/7-77-052, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, May 1977.
4. R.B. Fling et al., Disposal of Flue Gas Cleaning Wastes: EPA Shawnee Field Evaluation: Second Annual Report, EPA-600/7-78-024, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, February 1978.
5. J. Rossoff et al., Disposal of By-Products from Nonregenerable Flue Gas Desulfurization Systems: Final Report, EPA-600/7-79-046, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, February 1979.
6. R.B. Fling et al., Disposal of Flue Gas Cleaning Wastes: EPA Shawnee Field Evaluation: Third Annual Report, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina (In Publication).
7. J. Rossoff et al., Landfill and Ponding Concepts for FGD Sludge Disposal, The Aerospace Corporation, Presented at the U.S. Environmental Protection Agency Industry Briefing Conference, Raleigh, North Carolina, August 1978.
8. J. Rossoff et al., Landfill and Ponding Concepts for FGD Sludge Disposal, The Aerospace Corporation, Presented at the Air Pollution Control Association 71st Annual Meeting, Houston, Texas, June 1978.
9. J. Rossoff et al., Flue Gas Desulfurization Waste Disposal Field Study at the Shawnee Power Station, The Aerospace Corporation, Environment and Energy Conservation Division, Presented at the U.S. Environmental Protection Agency RTP Symposium on Flue Gas Desulfurization, Hollywood, Florida, November 1977.

DEMONSTRATION OF FORCED OXIDATION

AT

TVA'S WIDOWS CREEK UNIT 8

LIMESTONE FGD SYSTEM

By

R. A. Runyan

Environmental Control Technology - Air
Energy Demonstrations and Technology
Tennessee Valley Authority
Chattanooga, Tennessee

Preceding page blank

This preliminary critique entitled "Demonstration of Forced Oxidation at TVA's Widows Creek Unit 8 Limestone FGD System," has been prepared for presentation by TVA, Environmental Control Technology, at the Environmental Protection Agency Fifth Shawnee Industry Briefing Conference on December 5, 1979, conducted at the Royal Villa Motel in Raleigh, North Carolina.

The Forced Oxidation Demonstration Project primary test objectives were achieved by November 15, 1979. The test program has been in the operational phase the last eight months. Presently, the operational procedures and the test data are being evaluated and analyzed and will be presented in a detailed technical report to be published by TVA in the spring of 1980.

INTRODUCTION

TVA contracted with Combustion Engineering (CE) to install forced oxidation equipment on the limestone flue gas desulfurization (FGD) system at the Widows Creek Steam Plant and demonstrate that forced oxidation of FGD wastes is possible at this location as a processing scheme for waste disposal. The forced oxidation demonstration program began on April 2, 1979, on "D" module of the unit 8 scrubber and continued until November 15, 1979. Combustion Engineering had the responsibility for the initial operational phase through June 30 after which Radian Corporation assumed the operational phase responsibility through November 15, 1979.

The unit 8 limestone FGD system at Widows Creek consists of four parallel scrubber trains which process flue gas from a 550-MW CE system generator. The waste slurry produced by the FGD system currently is stored in a pond. Disposal of slurry represents a major problem in continued, long-term operation of the scrubber unit. It was decided to evaluate and demonstrate the forced oxidation method for treating the sludge to decrease the effective area required for disposal or to improve the stability of disposal areas. During the demonstration program, a forced oxidation system was installed on scrubber train "D." Approximately 10 percent of the system bleed (4370 #/hr of solids) of the oxidized scrubber bleed stream was processed through a 2-stage dewatering system consisting of a thickener and vacuum filter.

The demonstration of forced oxidation was divided into three separate experiments. These are as follows:

EXPERIMENT A*

Oxidation in the absorber effluent hold tank only. The objective of this experiment was to demonstrate that forced oxidation is possible in a single loop system.

EXPERIMENT B

Oxidation in the venturi effluent hold tank only. This experiment was conducted to test the hypothesis that the air stoichiometry required for oxidation in this tank will be significantly less than in the absorber tank because of lower pH.

EXPERIMENT C

Oxidation in both venturi and absorber effluent hold tanks. This experiment was conducted to determine if dual tank oxidation can be accomplished at a lower oxygen stoichiometry than single tank oxidation.

*Experiment A was deleted on the basis of limited funds and project schedule commitments.

Initially, the filter cake from the vacuum filter was reslurried and disposed of in the pond. After the test objectives had been achieved and conditions determined for producing a cake of consistent quality, arrangements were made to initiate the landfill disposal project.

PROGRAM OBJECTIVES

The general objectives of the program are:

- Demonstration of forced oxidation as a viable FGD waste disposal option.
- Acquisition of data applicable to the design of a forced oxidation system for the Widows Creek units 7 and 8 FGD system.

Three specifically quantitative goals of the demonstration program are:

- Attainment of greater than 95 mole percent conversion of sulfite to sulfate at commercially acceptable air stoichiometries.
- Production of a waste product capable of being dewatered to greater than 80 wt percent solids in a vacuum filter.
- SO_2 removal efficiency of 85 percent or greater.

FGD SYSTEM DESCRIPTION

A partial flow schematic of the Widows Creek unit 8 forced oxidation demonstration system is shown in Figure 1. Also indicated in Figure 1 is the coupling of the subscale dewatering train to the existing FGD system.

The pressurized scrubber system consists of four identical 25 percent capacity trains located downstream of low efficiency electrostatic precipitators. The principal components in each train include a 25 percent boiler I.D. fan, venturi scrubber, grid-type absorber, entrainment separator, flue gas reheat system, venturi slurry recirculation system, and absorber slurry recirculation system. Flue gas bypass capability is provided for each of the four scrubber trains.

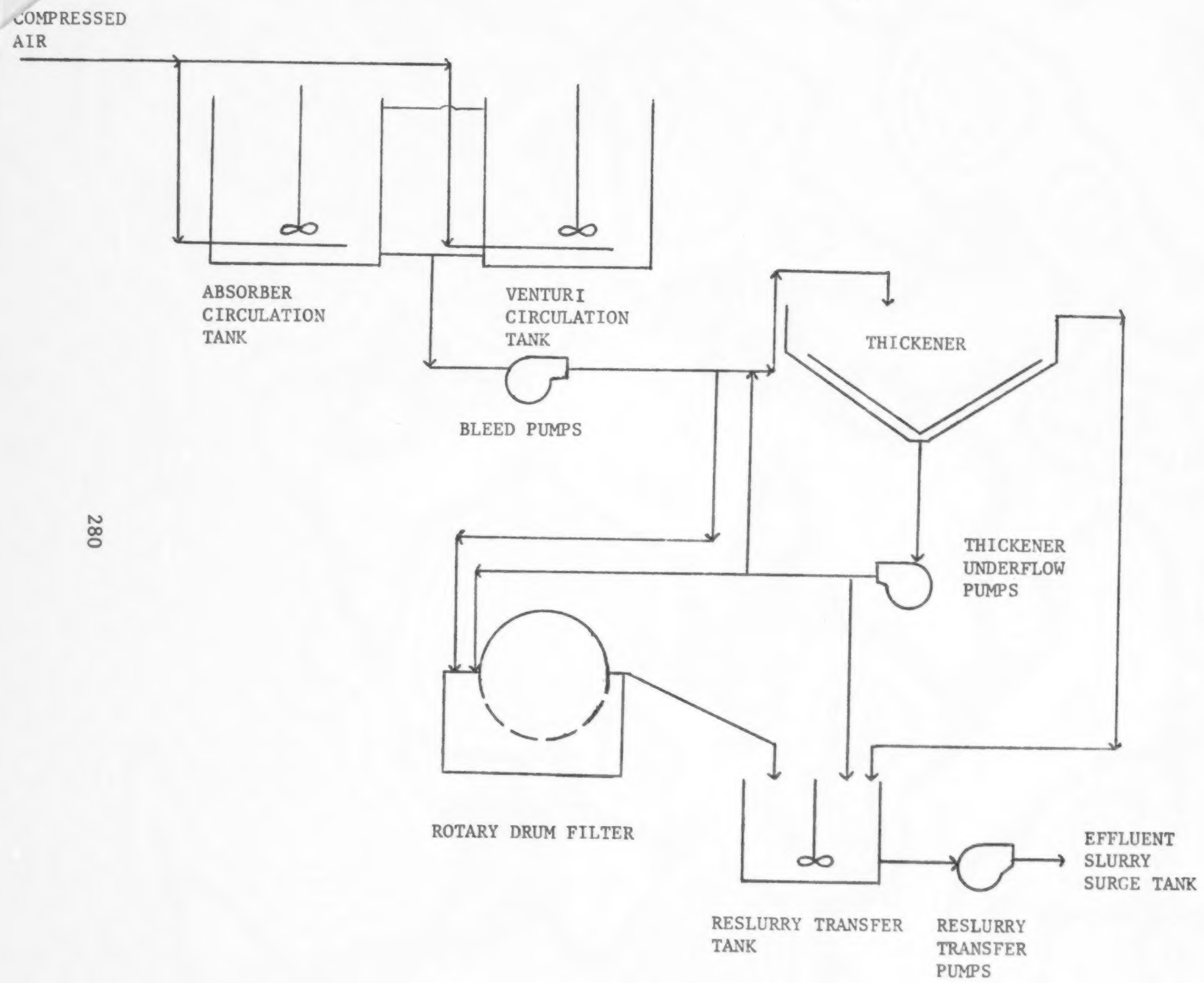

COMPRESSED AIR
ABSORBER CIRCULATION TANK
VENTURI CIRCULATION TANK
BLEED PUMPS
THICKENER
THICKENER UNDERFLOW PUMPS
ROTARY DRUM FILTER
RESLURRY TRANSFER TANK
RESLURRY TRANSFER PUMPS
EFFLUENT SLURRY SURGE TANK

FIGURE 1: FORCED OXIDATION DEMONSTRATION WIDOWS CREEK UNIT 8

PRINCIPAL TEST VARIABLES

There are two major goals associated with forced oxidation studies. The first goal pertains to our ability to "force" the chemical reaction of aqueous sulfur with oxygen. The second goal is related to the effects of the chemical oxidation reaction on the liquid-solid separation performance of thickener and vacuum filtration units.

OXIDATION

The prime requirement of this program is to demonstrate that aqueous phase sulfur can be oxidized from the (+4) to the (+6) oxidation state in the context of the Widows Creek FGD system. Essentially, this demands that the following reactions be thermodynamically favored and kinetically attainable at process conditions.

$$SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=}$$

$$HSO_3^{-} + 1/2\ O_2 \rightarrow SO_4^{=} + H^{+}$$

Sulfite-bisulfite oxidation is heterogeneous in nature since oxygen must diffuse from the gas to the liquid phase prior to chemical reaction. Thus, the overall oxidation rate can be a function of interphase oxygen mass transfer and/or chemical reaction kinetics. In a flue gas desulfurization system, the principal variables capable of influencing the oxidation rate are:

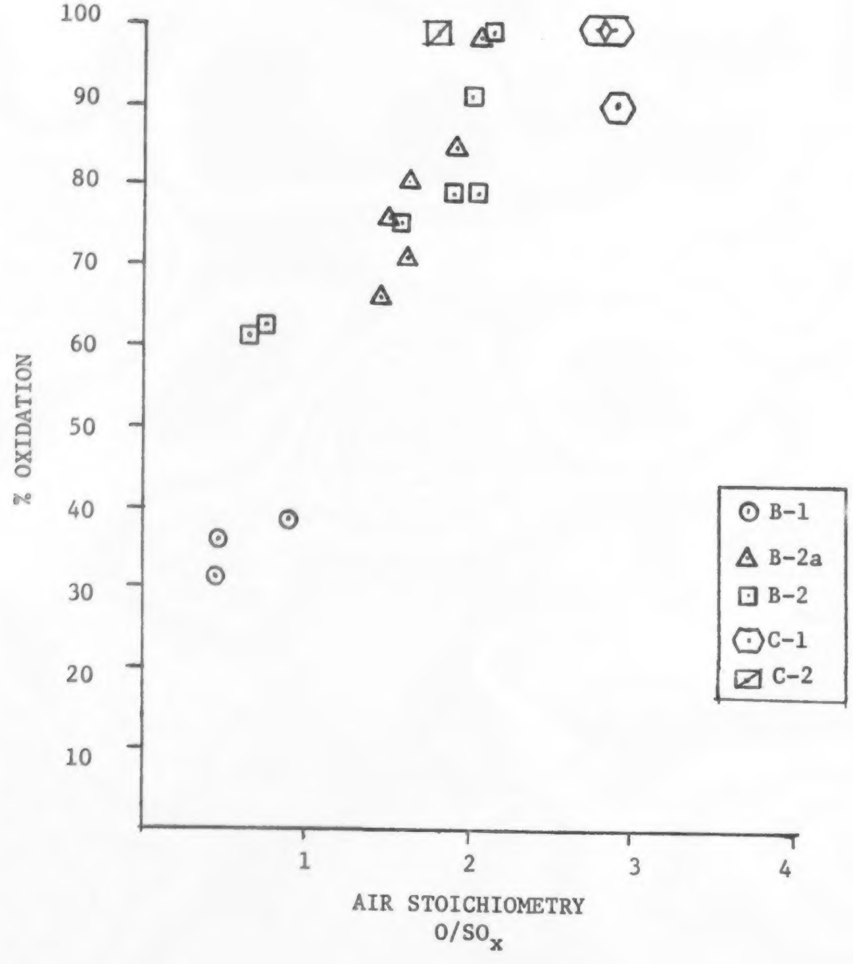
- Oxygen/sulfur stoichiometry
- Hold tank liquid level/residence time
- Slurry pH/limestone stoichiometry
- Slurry solids concentration
- Air sparger geometry
- Power input to mixer
- Solution ionic strength

SOLIDS DEWATERING

As a consequence of forced oxidation, it is hoped that dewatering properties of FGD waste solids can be significantly improved. It will be a specific intent of the program to demonstrate that a highly oxidized waste product can be dewatered to better than 80 percent solids in a thickener-vacuum filter train.

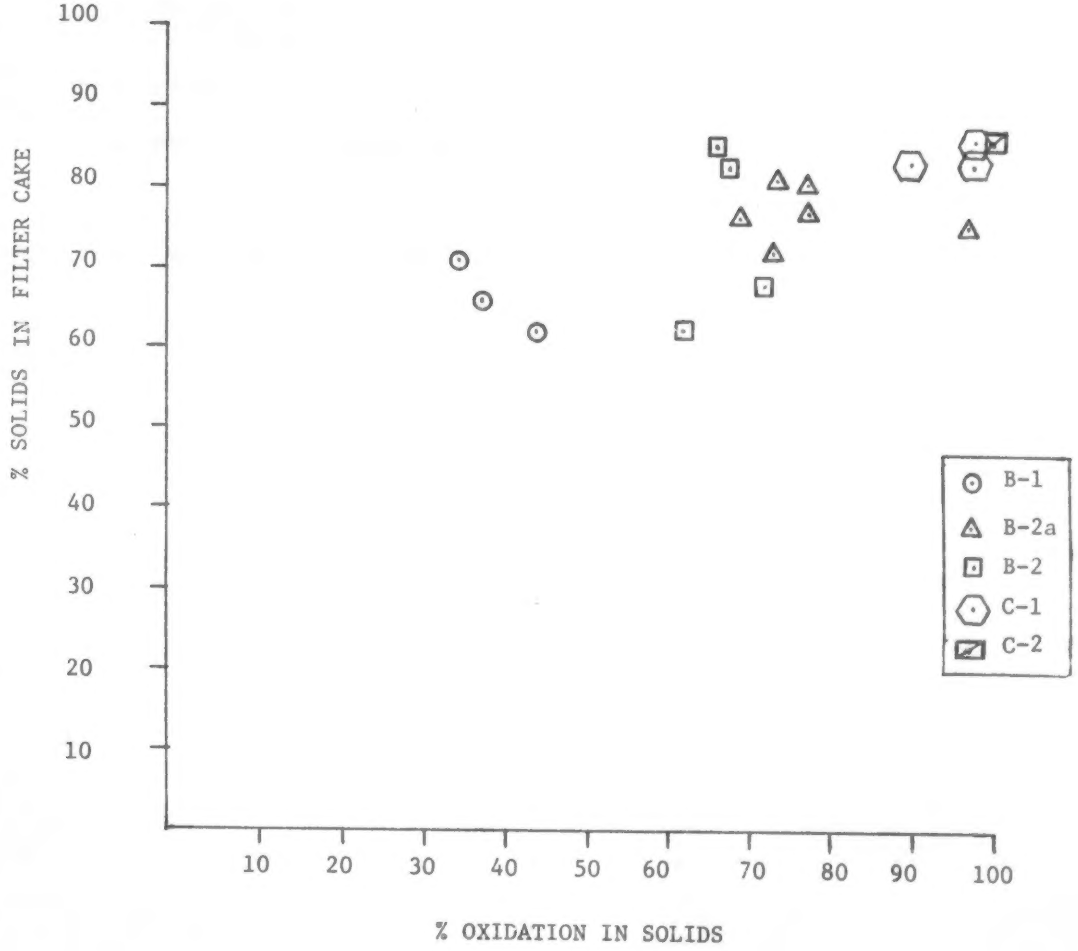
The performance of the dewatering equipment will be optimized as a function of the following operation variables.

- Thickener underflow recycle rate
- Flocculent addition rate
- Vacuum filter drum rotation speed
- Vacuum filter pressure differential
- Vacuum filter drum submergence depth



(LB ATOMS O / LB MOLE SO_2 ABSORBED)

% OXIDATION VS AIR STOICHIOMETRY



FILTER CAKE % SOLIDS VS % OXIDATION IN SOLIDS

RESULTS

- Percent oxidation (calcium sulfite to calcium sulfate): 95-99 percent
- Percent solids in the filter cake: 80-85 percent
- Percent SO_2 removal efficiency: 88-92 percent
- Air stoichiometry (lbatoms O/lbmole SO_2 absorbed): 1.8-2.0
- Limestone stoichiometry (lbmole $CaCO_3$/lb mole SO_2 absorbed): 1.2-1.4

SCRUBBER SYSTEM MODIFICATIONS

- Absorber circulation tank slurry overflow

  Recommends an overflow line between the absorber and venturi tanks at a 1-foot lower elevation than the original overflow pipe. This modification eliminated the slurry overflow problem at the top of the absorber tank while air was being injected into the tank.

- Absorber circulation tank downcomer

  The original downcomer position allowed an air lock in the downcomer, thus, restricting slurry flow. Extension of the downcomer below the air sparging ring in order to eliminate the air intake in the downcomer, thus, allowing slurry flow.

- Venturi and absorber circulation tank agitators

  Presently investigating the possibility of redesigning the impellers in order to achieve a 98 percent plus for forced oxidation.

- Absorber packing

  Packing placed on top of the second, third, fourth, and fifth grid levels in the absorber. The modification was necessary to achieve minimum SO_2 removal efficiency required in the test plan objectives.

END

12-18-80